Study of Reaction Kinetics of a Submerged Membrane Activated Sludge Process

by

Iqbal Basha Kalyandurg

A Thesis Presented to the

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In Partial Fulfillment of the
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In

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This dissertation, written by **Kalyandurg Iqbal Basha** under the direction of his dissertation advisor and approved by his dissertation committee, has been presented to and accepted by the Dean of Graduate Studies, in partial fulfillment of the requirements for the degree of **DOCTOR OF PHILOSOPHY IN CIVIL ENGINEERING**.

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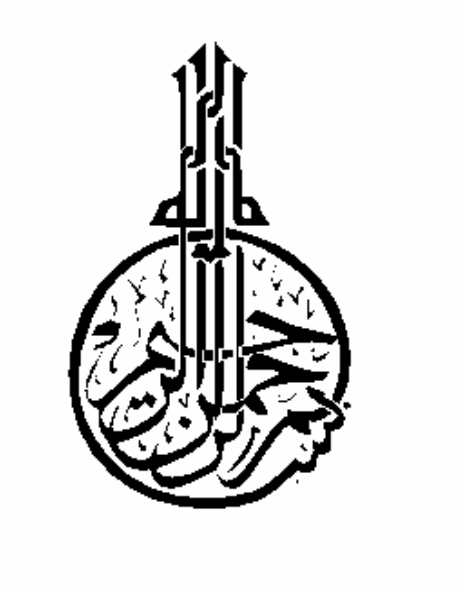

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DISSERTATION ABSTRACT

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The recent advancement in the membrane technology, especially in microfiltration, has given an impetus to the development of membrane bioreactors for the treatment of various wastewaters especially for the treatment of municipal wastewater. A submerged membrane activated sludge process (SM-AS) is one of the modifications to the conventional activated sludge process. It is the combination of a membrane module and a bioreactor. This SM-AS process can retain a high concentration of mixed liquor suspended solids (MLSS) in the aeration tank, giving numerous benefits over conventional wastewater treatment processes. Behavior of microorganisms in the SM-AS processes was needed to be understood so that these types of treatment plants could be designed to meet required effluent standards.

In this research a lab scale submerged membrane activated sludge process was operated for over one year period to study the kinetic coefficients of the SM-AS system under different MLSS (3000 mg/l, 5000 mg/l, 10 000 mg/l and 15 000 mg/l) concentrations. The Monod kinetic coefficients, Yield (Y) varied from 0.487 to 0.583 mg/mg; Endogenous decay coefficient (kd) varied from 0.151 to 0.0261 day-1; Maximum specific growth rate (μm) varied from 1.28 to 6.46 day-1 and Saturation constant (Ks) varied from 289 to 2933 mg COD/l. The study also identified the microorganisms present in the aeration tank. Fouling of the membrane was controlled by intermittent pumping schedule, back flushing with air and/or mechanical cleaning with a brush. The system could withstand easily shock loading of 16000 mg/l COD at an MLSS concentration of 15000 mg/l giving over 98 % COD removal efficiency. The system was also tested for the ability to withstand toxic loadings. The SM-AS system could withstand 400 mg/l phenolic toxic loading giving an effluent COD removal of 77%. With acclimatization with phenol, the effluent COD removal efficiency increased to 81%. The system could withstand chromium dosage of 50 mg/l and the system regaining its capacity within couple of days of removal of toxic loading.

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إن التّقدّم الحديث في تكنولوجيا الغشاء وخاصةً في الترشيح بالإنتشار الغشائي, قد أعطى دافعا لتطوير المفاعلات الحيوية الغشائية لمعالجة مياه الصّرف المتنوّعة خاصّة معالجة مياه الصّرف الصحي. إن عملية المفاعل الحيوي الغشائي المغمور هي أحد التّعديلات على عمليّة الحمأة المنشّطة التّقليديّة حيث تجمع بين الغش اء والمفاعل الحيوي. هذه العملية تتميز على عمليّات معالجة مياه الصّرف التّقليديّة بالعديد من الفوائد منها إمكانية الإحتفاظ بتركيز عالٍ من المواد الصلبة العالقة في الحمأة المنشّطة في خزان التهوية. وقد كان من الضروري فهم سلوك الكائنات الدّقيقة في هذه العملية حتّى يمكن أن تُصَمَّم هذه الأنواع لمحطّات المعالجة لمقابلة مقاييس المخلّفات السّائلة المطلوبة.
في هذا البحث تم تشغيل مفاعل حيوي غشائي مغمورذي حجم معملي لمدّة سنة لدراسة المعاملات الحركيّة لهذا النظام تحت قيم مختلفةلتركيز المواد الصلبة العالقة في الحم أة (3000ملغم/ لتر, 5000ملغم/ لتر, 10000 ملغم / لتر و15 000 ملغم / لتر) حيث تباينت معاملات مونود الحركيّة ما بين 0.487 إلى 0.583 في اليوم, وتباين معامل الخراب باطني النّموّ ما بين 0.151 إلى 0.0261 يوم. تباين معدّل النّموّ المعيّن الأقصى ما بين 1.28 إلى 6.46 يوم وتباين ثابت التّشبّع ما بين 289 إلى 2933 ملغم COD / لتر . تعرّفت الدّراسة أيضًا على الكائنات الدّقيقة الموجودة في خزان التهوية. تم التحكم بممانعة الغشاء بواسطة جدول ضخّ متقطّع, حيث تم الشطف بالهواء أو التّنظيف الميكانيكيّ بفرشاة. يمكن للنظام أن يقاوم بسهولة صدّمة تحميل مقدارها 16000 ملغم COD/ لتر في تركيز 15000 ملغم / لتر معطيا ما يزيد عن 98 % لفاعليّة إزالة المخلفات. أختبر النّظام أيضًا للقدرة على مقاومة التحميل السّام حيث قاوم تحميل فينولي سام مقداره 400 ملغم / لتر معطيا 77 % لفاعليّة إزالة المخلفات. بالتّكيّف بالفينول, زادت فاعليّة إزالة المخلفات إلى 81 % . تمكن النّظام من مقاومة جرعة الكروم البالغة 50 ملغم / لتر وإسترد النّظام سعته لإزالة التحميل السّام خلال أيّامٍ قلائل.

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CHAPTER 1

INTRODUCTION

For many years, the need for the retention of a high concentration of biomass in biological wastewater treatment systems has attracted the interest of environmental engineers and scientists. The potential benefits that could be obtained from such a high biomass concentration (Mixed Liquor Suspended Solids, MLSS) in an Activated Sludge system treating organic substrate may be listed as:

1. A high MLSS concentration in the reactor means small foot print of aeration tank
2. A high MLSS can adsorb and withstand shock loadings and to some extent toxic loadings
3. Lower surplus sludge production

The conventional activated sludge process commonly employs MLSS concentrations of 2000-3500 mg/l, not because these are optimal but because this is the maximum that can be achieved using gravity for solid/liquid separation in the secondary sedimentation tank. The final settling tank is a vital part of the activated sludge process. It combines two functions, clarification & thickening. Failing to provide either of these two functions

results in solids being carried over with the final effluent. This will not only affect the effluent quality, but can also affect the behavior of the biological process.

Membrane separation technology offers an attractive alternative for solid/liquid separation since it is possible to retain up to 100% of the biomass and thus run the aeration tank at a higher level of mixed liquor suspended solids. The use of a membrane for solids separation instead of a gravity clarifier eliminates many of the solids separation problems associated with conventional activated sludge process, such as low settling rates caused by high MLSS concentrations and filamentous bulking and other such as filamentous growth and pinpoint floc (Metcalf and Eddy, 1991).

There are two configurations for membrane bioreactors (MBR): External membrane filtration bioreactors and internal submerged MBRs. In the external MBR configuration, mixed liquor is pumped from an aeration tank to a pressure-driven membrane system outside the bioreactor where the suspended solids are retained and recycled back to the bioreactor and the effluent passes through the membrane. The membranes are regularly backwashed to remove suspended solids build-up and accumulations, and are chemically cleaned when the operating pressures become too high. Thus high quality effluent, independent of the MLSS concentration and characteristics of floc settliability can be achieved. However there is a possibility of breaking of microbial flocs due to high shear stresses induced because of high crossflow velocity, resulting in loss of viable microbial mass in the aeration tank. Also because of recirculation pump energy costs were higher.

The submerged membrane separation activated sludge process is gaining importance in municipal wastewater treatment. It alleviates the above problem of shear lysis of

microbes due to high crossflow velocity. In the submerged MBR configuration, a low-pressure membrane is submerged in the aeration tank and operated under vacuum pressure. The membrane is agitated by coarse bubble aeration that helps prevent suspended solids accumulation at the membrane surface. The submerged membranes are either regularly backwashed or relaxed, and are chemically cleaned when the operating pressures become too high. It requires no circulation pumps thereby making it an energy conserving system.

The use of the membranes allows the MBR's to operate independent of sludge settling qualities and eliminates the need for secondary clarification. As a result, a typical submerged MBR can operate at MLSS concentration between 5,000 to 40,000 mg/l (Yamamoto et al, 1989, Yamamoto and Win 1991 and Building Res. Inst. 1998). Also most submerged MBRs operate at sludge ages in excess of forty days. Thus, the operational parameters of submerged MBR would be different than that of conventional biological wastewater systems. Although much research has been done, since the first use of this technology by Yamamoto et al, 1989, they were related mostly to the organic removal efficiencies, and sludge retention times. The literature lacks information regarding the biokinetic coefficients of submerged membrane bioreactors.

Based on the above comments, it was considered necessary for an investigation to be carried out to study the performance of the submerged membrane activated sludge (SM-AS) system to determine the biokinetic coefficients and to verify a mathematical model for the submerged membrane activated sludge process. Also, it was felt necessary to investigate ability of the SM-AS system to withstand shock loading and toxic loadings.

CHAPTER 2

LITERATURE REVIEW

2.1 INTRODUCTION

Activated sludge processes, which treat both domestic and industrial waste waters, is currently the most widely used biological wastewater treatment process in the developed world. In this process, large amounts of biosolids are formed and the separation, dewatering, treatment and disposal of this sludge represent major investment and operating costs. Also, as many of the existing treatment plants operating on conventional activated sludge process have exceeded their design capacities. Therefore it becomes interesting to modify the activated sludge process in order to meet the challenges thrown.

Membrane bioreactor technology has been utilized in wastewater treatment as a modification of the conventional activated sludge process, where the separation of the effluent is facilitated by membrane filtration instead of sedimentation. The absolute retention of all microorganisms by a membrane makes it possible to treat wastewater effectively. The membrane bioreactor process has had full-scale application in a number of areas including industrial wastewater treatment, municipal wastewater treatment, landfill leachate treatment, domestic water reuse and drinking water reclamation

(Jefferson et al., 2000). The advantages of membrane bioreactor process are maintenance of high biomass concentration, reduced sludge production, high effluent quality and compactness.

2.2 CONVENTIONAL ACTIVATED SLUDGE PROCESSES

In general, activated sludge processes is a continuous or semi continuous (fill and draw) aerobic method for biological wastewater treatment. These processes are based on the aeration of wastewater with flocculating biological growth, followed by separation of treated wastewater from this growth. Part of this growth is then wasted, and the remainder is returned to the system. Usually, the separation of the growth from the treated wastewater is performed by settling (gravity separation) but it may also be done by flotation and other methods.

The activated sludge process presently represents the most widespread technology for wastewater treatment. Activated sludge plants can be found in different climate conditions - from the tropics to the polar regions, from sea level (wastewater treatment plants in ships) to extreme elevations (mountainous hotels). The scale of activated sludge plants ranges from package plants for one family to huge plants serving big metropolises. Wastewater treatment plants equipped with the activated sludge process are able to fulfill the most stringent effluent criteria.

2.2.1 Historical Evolution

The activated sludge process (ASP) for treatment of municipal wastewater was developed at the Davyhulme Treatment Works in Manchester, England from 1912 to 1914 by

Ardern and Locket (Ardern and Lockett, 1914). By means of discontinuous aeration experiments, it was discovered that wastewater cannot be treated exclusively by aeration from the contained pollutants, but that it must be activated with sludge and/or the living microorganisms contained in it. If the aeration of the wastewater and the sludge mixture was put down, the activated sludge settled in the form of flocs and the supernatant was free of the pollutants. Since then, the basic process has been widely adopted and further developed, giving it a unique flexibility of operation.

The activated sludge process surpasses all other biological procedures from a technical and economical point of view for purification of municipal wastewater treatment (Fair and Geyer, 1959). The success of the activated sludge process is due to the great performance of the overall system compared to the extremely variable process conditions while processing municipal wastewater.

2.2.2 Activated Sludge Process Description

Operationally, biological waste water treatment with activated sludge process is typically accomplished as shown in Figure2.1. Organic waste is introduced into an aeration tank, which contains a large population of microorganisms, where the substrate is utilized to yield more biomass and to produce energy needed for growth. After a specified period of time, the mixture of cells is passed into a settling tank, where the cells are separated from the treated wastewater. A proportion of the settled biomass is recycled to the inlet of the aeration tank to maintain the desired level of microorganisms in contact with organic waste. The remainder is wasted as concentrated sludge.

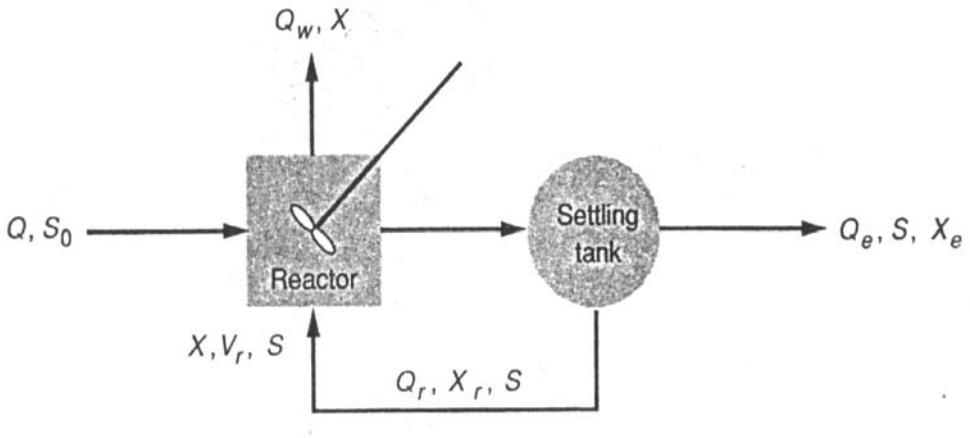

Figure 2.1 Schematic diagram of an activated sludge process

The oxygen is supplied, either by mechanical aeration or air blowing system. In mechanical aeration processes, surface aerators are utilized to provide oxygen by intensive agitation to produce extensive air-liquid interface. Provision of the oxygen to the biomass is controlled by altering the depth of immersion of the aerator either by raising/lowering the liquid level in the aeration tank with an adjustable outlet weir or less commonly, by a rise-and-fall gearbox fitted to the aerator. In the case of the air blowing system, compressed air is provided through air diffusers spaced evenly over the base of the tank. The air is intended to supply oxygen to the biomass as well as provide mixing. Typically, biomass concentration is varied between 2000 to 4000 mg/l. Because of the mixing pattern inside the tank, the oxygen supply is not efficiently utilized, since at the influent end more oxygen is needed than at the down stream end where most of organic substrate is removed.

2.2.3 Physical and Biochemical Reactions

The concentration of activated sludge in the reactor is indicated by the mixed-liquor suspended solids (MLSS). The total weight of MLSS (W_{ds}) is made up of the following constituents (Stewart, 1972):

$$W_{ds} = W_a + W_e + W_i \qquad (2.1)$$

where W_a = active biomass; W_e = inactive endogenously produced biomass; and

W_i = unassimilated portion, which is due to influent suspended solids that are either absorbed or remain unattached.

For the treatment of municipal wastewater, it is assumed that the biological processes are dominated by the growth of heterotrophic bacteria (Metcalf & Eddy 1991). The product of this biological process is designated as biomass. The average composition of biomass and activated sludge is indicated in Table 2.1.

Metabolic reactions consist of assimilation and breakdown of protoplasm into elemental constituents. These reactions can be described by the following stoichiometric equations (Stewart, 1972):

Energy Production or Respiration

$$\text{Organic Matter} + O_2 \xrightarrow{\text{Microorganisms}} CO_2 + H_2O + \text{Energy} \qquad (2.2)$$

Protoplasm Synthesis

$$\text{Organic Matter} + NH_3 + P \xrightarrow{\text{Microorganisms, Energy}} \text{Protoplasm (living cells)} \qquad (2.3)$$

Endogenous Respiration or Cell maintenance

$$\text{Protoplasm} + O_2 \rightarrow CO_2 + NH_3 + H_2O + \text{Dead Cells} \qquad (2.4)$$

The gross weight of biomass, net weight of biomass and BOD removal all vary with the ratio of daily weight of BOD applied to the weight of activated sludge, i.e. organic loading rate (OLR) or the food to microorganisms (F/M) ratio (Al-Layla et al., 1980).

Table 2.1 Average Composition of Biomass and Activated Sludge

Constituent	**Symbol**	**Biomass***	**Activated Sludge****
Organic Carbon	C	50%	40 – 45%
Hydrogen	H	8%	--
Oxygen	O	20%	--
Nitrogen	N	14%	8 – 10%
Phosphorus	P	3%	2 – 2.5%
Loss in ignition	--	80 – 90%	62 – 92%

* Schlegel, 1985 and Uhlmann, 1988

** Popel, 1973

2.3 MICROBIAL KINETICS

In biological wastewater treatment the most widely occurring and abundant group of microorganisms are the bacteria, and it is these groups which are most important in terms of utilizing the organic matter (the substrate) present in the wastewater. Research into the kinetics of microbial growth has proceeded down two paths during the last fifty years. One path has been fundamental in scope and has been followed by experimenters using pure microbial cultures. The other has been more applied and has been followed by researchers working with mixed microbial cultures. Both paths have their starting points in the early work of Monod (1949).

2.3.1 Enzyme kinetics

The rates at which microorganisms carry out their various functions depend upon the interrelationships between the rates of energy supply and energy utilization, which are determined by the rates of the enzymes in the pathways. The first generally acceptable mathematical formulation for the kinetics of enzyme reactions was put forward by Michaelis and Menten and subsequently served as the foundation upon which most current formulas are based. The basic assumption is that enzymatic catalysis occurs through a series of elementary reactions involving the enzyme-substrate complex. This can be shown as:

$$C_S + C_E \underset{k_{-1}}{\overset{k_1}{\rightleftharpoons}} C_{ES} \overset{k_2}{\Rightarrow} C_E + C_P \qquad (2.5)$$

where C_S, C_E, C_{ES} and C_P represent the concentration of the substrate, free enzyme, enzyme-substrate complex and product respectively. Michaelis-Menten equation can be written as

$$v = -r_s = V_m C_S / (K_m + C_S) \quad (2.6)$$

where,

v = velocity of an enzymatic reaction

$-r_S$ = rate of consumption of the substrate

$$V_m = k_2 C_{E0} \quad (2.7)$$

C_{EO} = initial enzyme concentration

K_m is called "Michaelis-Menten constant"

$$K_m = (k_{-1} + k_2) / k_1 \quad (2.8)$$

2.3.2 Biochemical Kinetics

The growth of a microbial culture is a complex phenomenon composed of a number of simultaneously occurring events. They can be grouped into the following three categories:

- cell growth and substrate utilization,
- microbial death and viability, and
- microbial decay.

The relative magnitudes of the respective rates determine what the net effect is upon the culture. Experimentally, it has been found that the effect of a limiting substrate or nutrient

can often be defined adequately using the following expression proposed by Monod (1949):

$$\mu = \frac{\mu_m S}{K_s + S} \tag{2.9}$$

where μ = specific growth rate, day^{-1}

μ_m = maximum specific growth rate, day^{-1}

S = concentration of growth limiting substrate in solution, mg/l

K_s = half-velocity constant, substrate concentration at one-half the maximum growth rate, mg/l

The use of mathematical models for the simulation of wastewater treatment processes has gained widespread acceptance as a tool to aid the design of new works (Diagger & Nolasco, 1995), and optimization of existing facilities (Chambers & Jones, 1988; Barnet and Sedaraty, 1994; Coen et al., 1997 and Horn & Chen, 1997). There are several models of the activated sludge process. These can be separated into those based on the 5-day biochemical oxygen demand (BOD_5) as the measure of organic strength and those based on the chemical oxygen demand (COD) (Stokes et al., 2000).

The development of appropriate kinetic models for these systems is based on the following assumptions:

i. Complete mixing is achieved in the aeration tank,
ii. Influent substrate concentration remains constant,
iii. No microbial solids are contained in the raw wastewater to the aeration tank,
iv. No microbial activity occurs in the secondary clarifier,

v. No sludge accumulates in the secondary clarifier and a reasonable efficiency of solids-liquid separation is accomplished,

vi. All biodegradable substrate is in the soluble form, and

vii. Steady state conditions prevail throughout the system.

For the system shown in Figure 2.1, the mean cell residence time θ_c, can be defined based on aeration tank volume, by the following expression (Metcalf & Eddy, 1991),

$$\theta_c = \frac{V_r \quad X}{Q_W X_W \quad + \quad Q_e X_e} \tag{2.10}$$

where θ_c = Mean cell residence time, days

V_r = Volume of the reactor, l

X = Concentration of volatile suspended solids in the reactor, mg/l

Q_w = Waste sludge flow rate, l/d

X_w = Concentration of volatile suspended solids in the waste

Q_e = Effluent flow rate, l/d

X_e = Concentration of volatile suspended solids in the effluent, mg/l

From the mass balance for the microorganisms in the entire system gives,

$$X = \frac{\theta_c \; Y(S_i - S_e)}{\theta(1 + k_d \, \theta_c)} \tag{2.11}$$

where Y = Maximum yield coefficient, mg/mg

S_i = Influent substrate concentration, mg/l

S_e = Effluent substrate concentration, mg/l

θ = Hydraulic detention time, days

k = Maximum substrate utilization per unit mass of organisms, mg/mg

k_d = Endogenous decay coefficient, $time^{-1}$

Performing the substrate balance,

$$S_e = \frac{K_s(1+\theta_c k_d)}{\theta_c(Yk - k_d) - 1} \tag{2.12}$$

where Ks = Half velocity constant, substrate concentration at one-half the maximum growth rate, mg/l

The kinetic model for the shock loadings in an activated sludge process was originally proposed by Ramanathan and Gaudy (1971) and later modified by Manickan and Gaudy (1982). The cell and substrate material balance equations written around the aeration tank are as follows:

$$\frac{V\,dx}{dt} = \alpha F X_R + \mu X V - k_d X V - F(1+\alpha)X \tag{2.13}$$

$$\frac{V\,ds}{dt} = F S_i - F(1+\alpha)S_e - \mu\frac{XV}{Y_t} \tag{2.14}$$

Under steady state conditions, these material balance equations are reduced to

$$X = \frac{Y_t[S_i - (1+\alpha) + \alpha X_R}{1+\alpha + (k_d/D)} \tag{2.15}$$

$$S_e = \frac{-b \pm \sqrt{b^2 - 4ac}}{2a} \tag{2.16}$$

$$a = \mu_{\max} - (1+\alpha)D - k_d \tag{2.17}$$

$$b = D\left|S_i - (1+\alpha)K_s\right| - \frac{\mu_{\max}}{1+\alpha}\left[S_i + \frac{\alpha X R}{Y_t}\right] + k_d\left[\frac{S_i}{1+\alpha} - K_S\right] \tag{2.18}$$

$$c = K_s S_i\left[D + \frac{k_d}{1+\alpha}\right] \tag{2.19}$$

$$X_W = VX\mu_n \tag{2.20}$$

where, α = Recycle ratio

D = Dilution rate, time^{-1}

X_R= Concentration of recycle sludge, mg/l

μ_{max}= Maximum specific growth rate, time^{-1}

Y_t= True cell yield, mg/mg

Y_o= Observed Yield, mg/mg

μ_n= Net specific growth rate, time^{-1}

The model was felt to be more in tune with the prevalent conditions and practice for activated sludge process, while remaining consistent with the theory of continuous culture. Also, it includes all the known factors that affect the effluent quality and its control, S_i, k_s, μ_{max}, k_d, Yt, X_R and α.

The effects of SRT on biomass flocculation and in turn on its settling characteristics have been the subjects of many researchers. It has been found that bio flocculation occurs only at a SRT above 2 days and below 15 days. Curds et al. (1968) postulated that it was due to the presence of a polymeric substance excreted by protozoa into the medium. At low SRT, protozoa will be washed away and as a result, dispersed non-flocculent bacterial will be dominant. Pavoni et al. (1972) attributed bio flocculation to the excretion of polymeric substance by the bacteria themselves. The influence of SRTs longer than 15 days, lie in the increases in biomass decay rates (Ford and Ecnefelder, 1967). However, Harris and Mitchell (1975) related the deterioration in settling characteristics to the

appearance of small floc particles, called a pinpoint floc. This may have resulted from an excess of biopolymer.

Although quiescent sedimentation of biological solids is economical, the increase of public awareness concerning the environmental pollution is forcing the pollution control agencies to adopt other positive ways of controlling pollution. Crossflow membrane filtration gave an attractive alternative for solid/liquid separation. It had the added advantage of consistently producing an effluent almost free from suspended solids with less operational problems (Anderson et al., 1986; Vera et al., 1998 and Al-Malack et al., 1998). Moreover, the flocculating characteristics of the activated sludge are not relevant to the quality of the effluent, Bemberis et al., (1971), Arika et al., (1977) and Li et al., (1984).

2.4 MEMBRANE FILTRATION

Membrane filtration process is a pressure-driven separation of the components of a fluid mixture by selective permeation through an interphase (the membrane) separating the concentrate (or retenate) stream from the permeate stream. Permeate is defined as the stream that emerges from the membrane and which is depleted in one or more components. The concentrate (retenate) stream is defined as the stream on the upstream side of the membrane: this stream is enriched in the same components (Gutman, 1987).

The study of membranes began with the investigation of transportation processes through natural membranes such as animal intestines or fish air bladders, which led in the middle of the eighteenth century to the discovery of osmosis. In the middle of the nineteenth century the first synthetic membranes on a cellulose basis were produced. The basic laws

were set up in parallel by Hagen and Poiseuille for the flow in pores and by Fick for the diffusion processes. The industrial production and application of synthetic membranes on the basis of cellulose began around 1920. The evolution of polymer membranes developed around 1960 (Staude, 1992).

In general the driving forces can be generated by:

i. a concentration difference across the membrane,

ii. a pressure difference across the membrane,

iii. an electrical potential difference across the membrane, and

iv. a temperature difference across the membrane.

In a membrane separation process, more than one of these differences can be present together. However, for most membrane processes one particular component generally dominates the overall driving force, and the process is described as being driven by that component. Table 2.2 lists most of the currently identified membrane processes as well as their principal driving forces.

Table 2.2 Classification of membrane processes according to driving force (Gutman, 1987)

Driving force			
Hydrostatic Pressure	**Concentration difference**	**Electrical potential gradient**	**Temperature difference**
Microfiltration	Dialysis	Electrodialysis	Transport depletion
Ultrafiltration	Controlled release	Electro-osmosis	Thermo-osmosis
Reverse osmosis	Pervaporation	Electrophoresis	Membrane distillation
Nanofiltration	Gas separation		

2.4.1 Pressure driven membrane processes

The pressure driven membrane processes, micro-, ultra-, nanofiltration and reverse osmosis, are subdivided mainly accordingly to the cut-off of the inserted membrane. The cutoff of a membrane characterizes approximately the size of the particles and/or the molar mass of molecules, those that can still be rejected by the membrane. For cutoffs smaller than 0.1µm, the rejected components are mainly featured through their molecular weight in g/mol and no longer through their geometrical size. A direct correlation between the cutoff in µm and g/mol is not possible because of different molecule sizes at identical weights. However, it can be estimated that at a cutoff of 0.1µm, molecules with a molecular weight of more than 500,000g/mol will be rejected (Staude, 1992).

2.4.2 Membrane categorization and use

Membranes are categorized according to the size, number and distribution of their pores and the size of particles they can retain. In Figure 2.2 the pressure driven membrane processes from microfiltration to reverse osmosis are specified with the respective cutoffs and the typical range of application. The particle sizes of different matters are also given for the arrangement of the processes. Each utilizes a different separation mechanism and

has specific advantages and disadvantages when compared to the others for a particular application (Ferugusson, 1986). Table 2.3 gives comparison of various pressure driven membrane processes.

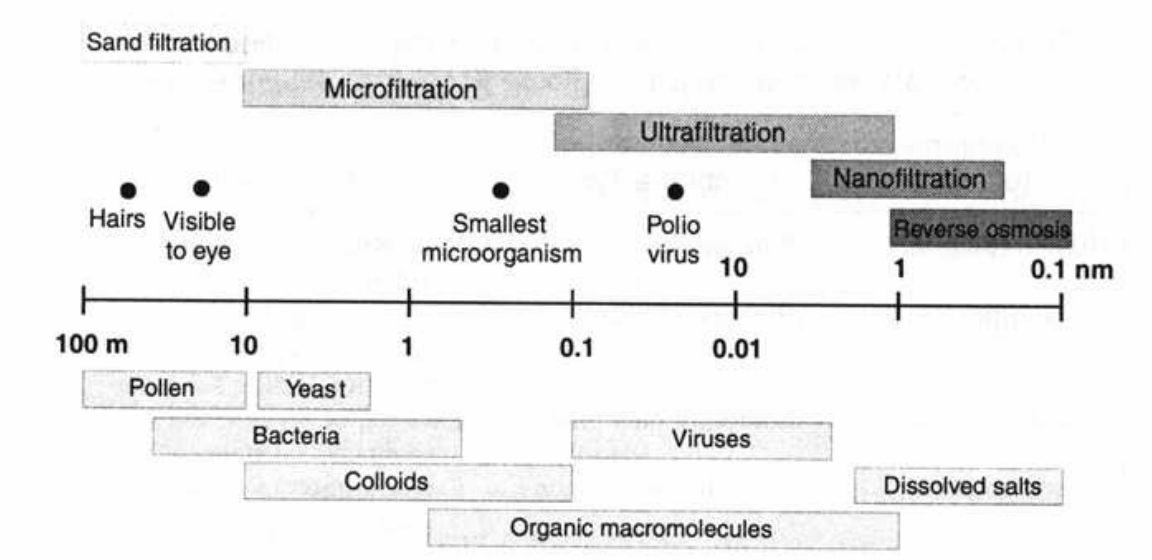

Figure 2.2: Classification of Pressure-Driven Membrane Processes and Particle Sizes.

(Gunder, 2000)

Table 2.3: Comparison of Various Pressure Driven Membrane Processes (Mulder, 1991).

MICROFILTRATION	ULTRAFILTRATION	REVERSE OSMOSIS
Separation of particle (Bacteria, Yeast)	Separation of macromolecules	Separation of low Molecular Weight solutes
Osmotic pressure is negligible	Osmotic pressure is negligible	Osmotic pressure is high (5 – 25 bar)
Applied Pressure is low (< 2 bar)	Applied pressure is low (1 – 10 bar)	Applied pressure is high (10 – 60 bar)
Symmetric structure (Not Always)	Asymmetric structure	Asymmetric structure
Thickness of separating layer = 10 – 150 μm	Thickness of separating layer = 0.1 – 1.0 μm	Thickness of separating layer = 0.1 – 1.0 μm
Separation is based on particle size	Separation is based on particle size	It is based on difference in solubility and diffusivity

The separation involved in Microfiltration can deal with removal of particulate or suspended materials ranging in size from 0.1 to 10 μm (Figure 2.2). On the other hand, Ultrafiltration is usually used to recover macromolecules in the 0.001 to 0.1 μm range. Reverse osmosis membranes are capable of separating materials less than 0.001 μm. Salt ion retention is a typical use of reverse osmosis (desalination of sea water). The operation of RO requires very high pressures sometimes as high as 150 bar in order to overcome the osmotic pressure, whereas, the hydrodynamic pressures required to induce flow through microfiltration and ultrafiltration membranes are generally in the region of 0.1 to 10 bar.

2.5 CROSSFLOW MICROFILTRATION

The aim of microfiltration is primarily the separation of particles from liquids. The cutoffs of the porous membranes used for this purpose are normally between 0.1 and 10 μm. However, membranes with cutoffs between 0.02 and 20 μm are also referred to in microfiltration, depending on the kind of solid matters to be separated and the aim of the membrane process (Ripperger, 1988). In the case of microfiltration, the particle rejection is based mainly on a sieve mechanism. The particles are restrained primarily due to their geometrical measurement on the membrane.

2.5.1 Procedures for Microfiltration

In the case of microfiltration, two procedures can be distinguished: dead-end filtration and crossflow filtration

Crossflow (also called tangential flow) filtration is the pressurized flow of the feed water or influent, across a membrane, with a portion of the feed permeating the membrane and

the balance of the feed sweeping tangentially along the membrane to exit the system without being filtered. The filtered stream is called “permeate”, because it has permeated the membrane. The second stream is called the “concentrate or reject”, because it carries off the concentrated contaminants rejected by the membrane. Because the feed and the concentrate flow parallel to the membrane instead of perpendicular to it, the process is called ‘crossflow’ or ‘tangential flow’.

In contrast, the so called ‘dead-end’ filtration is that where the flow direction is identical to the filtration direction and perpendicular to the filter medium, and within a short time filtered cake build up and filter blocking occurs. As shown in Figure 2.3, with the crossflow filtration it is intended to eliminate or minimize the cake from being built up by creating a shearing force mainly by flow at high velocity tangentially across the surface of the membrane. Dead-end filtration is generally not appropriate if the solution contains suspended solids in concentrations of more than 0.1% by weight. It is also unsuitable for filtration of suspensions, which contain colloidal material, or for production of very high quality effluent. Another disadvantage of dead-end filtration is that the separated solids cannot be utilized on a continuous basis, where as in crossflow filtration the recycled solids are easily attainable and readily utilized.

Crossflow filtration is gaining more acceptances as a useful separation process for the pharmaceutical, biological, water and wastewater treatment industries, since it can offer a complete removal of very fine colloidal solid material without addition of flocculants or filter aid. Further more, it can go down to molecular and ion level depending on the nature of the membrane used.

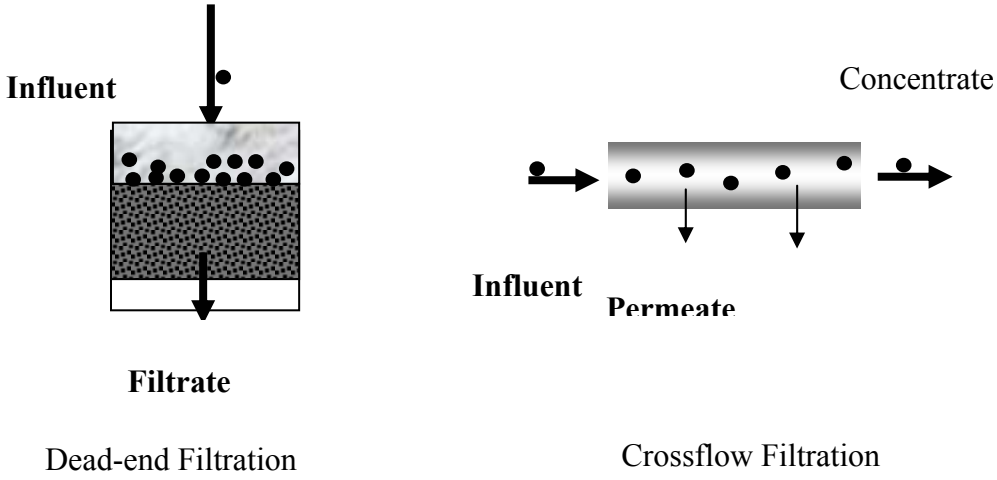

Figure 2.3 Comparison between Conventional and Crossflow Filtration

2.5.2 Crossflow Membrane Configurations

Generally there are two basic types of modules for crossflow membrane filtration, tubular and plane sheets membranes. Tubular modules are being made in hollow fibers, tubes and pleated modules. Plane or flat sheet membrane modules have been manufactured either in spiral wound cartridges or supported on frame holder.

Hollow Fiber Module

Hollow fibers having an inside diameter of less than 1.0mm, are aligned in a parallel fashion and are joined together on either end inside the housing vessel. The feed enters the housing at one end and flows through the interior of the membrane fibers. The reject material continues to flow through the fibers and is removed at the other end of the housing, while the permeate flows perpendicularly through the skin of the membrane. Hollow fiber membranes have an advantage of having a high packing density (high membrane area/volume ratio). Hollow fiber modules have the disadvantage of being highly susceptible to fouling. Their application is limited only to RO desalination plants and the polishing of the effluent of wastewater treatment.

Tubes and Pleated Modules

Pleated modules represent a new development and modification of tube membrane, with the added advantages: increased membrane surface area and strength. Tube membranes are more highly developed. They have a wide application in the field of fermentation process as these are less susceptible to blockages, and are also the simplest.

Flat Modules

These generally consist of sheets of membrane covering pores, support plates. As the feed flows between the membrane, the permeate leaves via the support plates and the concentrated feed continues to flow across the module for further treatment.

Spiral Wound Modules

The spiral wound configuration consists of flat sheet membranes, which cover each side of a flat surface of porous material with spacer screens separating the membranes during assembly. Porous material and spacer are wrapped around a perforated tube and form a spiral configuration. The perforated tube collects the filtered effluent. The spiral wound unit is subsequently inserted into a housing vessel. The feed is introduced between the spacer screen and flows through the membrane into the porous material and finally to the perforated tube.

The choice of a particular module for a particular application is dependent on a number of factors: cost of manufacture, ease of sampling and operating, energy efficiency and susceptibility to fouling. Each module has its advantages to offer along with its drawbacks. Table 2.4 summarizes the important features of these modules.

2.5.3 Microporous Membrane Formation

Microporous membranes for different type of membrane filtration have been prepared from a variety of materials including polymers, ceramics, metals and glass (see Table 2.5). Various ways and methods are being in use, depending on the materials, membrane and support structures, pore size, porosity and membrane thickness. There are four main basic structures that can be produced:

Table 2.4 Advantages and disadvantages of membrane modules (Harper 1980, Belter et al.1988, MacNeil and McCoy 1988)

Modules	Advantages	Disadvantages
Hollow fiber	Compact, packing density high Low hold up volume Economical	Very susceptible to plugging Difficult to clean Replacement of single tube is impossible
Tubular	High tolerance to suspended solids Easily cleaned chemically or mechanically Individual tubes can be replaced No dead spaces Well developed equipment	High hold up volume per unit area High pressure drop in tube connection Relatively expensive
Plate and frame	Low hold up volume Tolerance to suspended solids is high Well developed equipment	Difficult to design free of dead spaces Difficult to clean Expensive Entire module must be replaced on failure
Spiral wound	Compact, high packing density Relatively less expansive Fair tolerance to solids	Difficult to clean Entire module must be replaced on failure

Table 2.5 Organic and Inorganic Materials Used in Manufacturing Membranes (Al Malack, 1993)

Hydrophobic polymeric membranes	Polytetrafluoroethylene (PTFE, Teflon) Poly vinylidene fluoride (PVDF) Polypropylene (PP)
Hydrophilic polymeric membranes	Cellulose esters Polycarbonate (PC) Polysulfone/polyethersulfone (PSf/ PES) Polyimide/poly ether imide (PI/PEI) Aliphatic polyamide (PA)
Ceramic membranes	Alumina (Al_2O_3) Zincoria (ZrO_2)
Glass membranes	SiO_2
Metallic membranes	Palladium Tungsten silver

- Asymmetric membrane with uniform pore substructure
- Asymmetric membrane with a graded pore substructure
- Asymmetric membrane with a finger-pore substructure
- Symmetrical membrane with no skin layer

The most versatile and most widely used membrane preparation technique involves thermally induced phase separation. It consists of the following steps:

- A homogeneous solution is formed by melt-blending the polymer with high boiling, low molecular weight liquid or solid non-polymer. The non-polymer component of the solution is referred to as the 'diluent'
- Then the solution is cast into the desired shape
- The cast solution is cooled to activate phase separation

The diluent is removed by solvent extraction to produce a microporous structure

2.5.4 Fouling Theory in Crossflow Microfiltration

It is recognized that the highest obstacle facing the widespread use of membranes in the field water and wastewater testament industry is associated with the decline of the flux due to losses in membrane permeability. Flux decay can be a serious problem, which is usually a direct result of the formation of a so-called dynamic membrane or secondary membrane on the top of the primary membrane. The dynamic membrane is always formed either automatically or made through choice. This loss in permeability has been the subject of investigation and research for many years and several models for it have been proposed. Resistance to the filtration flux by the membrane material and particles in

the feed, and concentration polarization are the two main phenomena governing the flux rate decline in crossflow filtration.

Resistance Phenomenon

Darcy's Law which relates the flow rate (the flux) through a porous medium to the pressure drop can be applied. Darcy's Law states that the flux is directly proportional to the potential pressure drop and inversely proportional to the resistance (l/k). The resistance explicitly includes the contribution of cake and filter medium:

$$v = \frac{k\Delta P}{\mu l} \tag{2.21}$$

$$l/k = R_m + R_c \tag{2.22}$$

where

R_m = resistance caused by filter media; R_c = resistance caused by cake

Thus, when filtering a suspension containing a wide range of particles and colloids using a microporous membrane at a constant pressure, the filtration flux, J, can be expressed by the following resistance equation:

$$J = \frac{\Delta P}{\mu R_t} \tag{2.23}$$

where

R_t = total resistance to the flux = $R_m + R_c$

Concentration Polarization Phenomenon

When a solution flows through a membrane-bounded channel where the solvent passes through the membrane while the solute is rejected at the solution/membrane boundary,

the criterion for local steady state mass transfer of solute requires that the rate of convective transport of solute toward the membrane surface be equal to the rate of transport (by convective and diffusive mechanism) of solute away from the membrane surface (Michaels et al., 1987). This condition can be satisfied only if the solute concentration in the layer of solution adjacent to the membrane surface is higher than that in the bulk of the solution within the channel. In the simplest terms, concentration-polarization is the accumulation, at the upstream surface of the membrane, of solute molecules, which are rejected or retained by the membrane in the course of ultra- and microfiltration. Several investigators reported details about the concentration polarization concept (Bian et al. 2000 and Zhang and Song 2000).

Basically, the steady state mass transfer conditions which must exist within the polarized boundary layer are represented by the following relationship:

$$JC_x - D_s \frac{dC_x}{dx} - JC_p = 0 \tag{2.24}$$

where

J = transmembrane solvent flux; C_X = concentration at the boundary layer;

D_S = molecular diffusion coefficient; C_p = permeate concentration

By integrating, rearranging and assuming Cp=0 the above relation can be simplified to:

$$J = K_m \times \ln \frac{C_g}{C_b} \tag{2.25}$$

where

K_m = the particle mass-transfer coefficient between bulk suspension and membrane surface;

C_g = gel layer concentration; C_g = bulk concentration

2.6 MEMBRANE BIOREACTORS

2.6.1 Introduction

The combination of an activated sludge reactor and a membrane filtration unit for the separation of activated sludge is defined as membrane coupled activated sludge process or simply membrane bioreactors. Consequently, the difference from the conventional activated sludge process lies only in the separation of the activated sludge. In conventional final clarifiers only the fraction of the activated sludge that forms flocks and settles can be retained. With membrane filtration, all parts of the activated sludge that are larger than the cutoff of the membrane are retained. As a result, the separation of the activated sludge from cleaned wastewater is independent of the sedimentation qualities of the activated sludge and is only dependent on the microfiltration membrane.

To retain the bacteria contained in the activated sludge, which is necessary for the functioning of the process, microfiltraion membranes with a maximum pore size of 0.4 µm are usually used in the membrane bioreactors. To prevent fast clogging of the membranes, crossflow filtration is the only suitable filtration procedure for solid concentration of 3 000 mg/l and more as usually applied for the activated sludge process. In the following, the terms “membrane filtration,” “microfiltration” and “crossflow microfiltration” are used synonymously in connection with the membrane bioreactors. The continuous velocity on the membrane and/or the existence of turbulence near the membrane surface is defined as crossflow.

Application of membrane technology has begun to find its way in solid/liquid separation through development of a wider range of membrane structures. Membrane separation and retention of biological solids has been applied as one of the alternatives to conventional activated sludge process since late 1960's (Jae-Seok Kim *et al*., 2001). There are around 500 commercial membrane bioreactors in operation worldwide, with many more proposed or currently under construction (Stephenson *et al*., 2000). The boom in membrane process still continues today with an estimated growth in the market at the rate of 10 % per year, as new products and applications continue to emerge. The size of world pressure driven membrane market is now estimated to exceed £ 350 million per year. Of this total about £70 M is accounted for by sales of reverse osmosis (RO) equipment, £50 M is for sales of ultrafiltration (UF) equipment and £230 M for sales of microfiltration (MF) equipment.

The membrane bioreactors have shown many advantages, such as stable effluent quality, high volumetric loading and lower surplus sludge production (Engelhardt et al., 1998). Hence, studies on membrane bioreactors have attracted great attention for the treatment of domestic and industrial wastewater (Bailey et al., 1994; Chiemchaisri et al., 1993; Muller et al. 1995; Brindle and Stephenson, 1996; Ueda et al, 1996; Davies et al., 1998 and Wagner and Rosenwinkel 2000).

2.6.2 Types of Membrane Bioreactors

The membrane bioreactors consist of an activated sludge tank and a crossflow microfiltration unit for the separation of the activated sludge. Two variants can be distinguished with regard to the arrangements of the crossflow filtration unit:

1) Membrane bioreactors with external membrane filtration: The membrane filtration for separation of the activated sludge from cleaned wastewater is carried out outside the activated sludge tank. The membranes are put into corresponding modules, which must be with mixed liquor. The concentrate, that is the retained activated sludge, is returned to the activated sludge tank as return sludge.
2) Membrane bioreactors with internal submerged membrane filtration: The membrane filtration is carried out directly in the activated sludge tank. The membranes and/or suitable membrane modules are submerged in the tank. Therefore, a supply of the activated sludge and a recycling of the concentrate are not necessary. In practice, it is common to put the submerged membranes in a separate activated sludge tank, also called a filtration tank.

2.6.3 Historical Evolution of MBR

For the treatment of municipal wastewater normally the activated sludge process is used. In the aeration tanks conditions are created that stimulate the growth of various species of microorganisms, so that the pollutants are taken out and converted into cell material. The mixture of activated sludge and water is a suspension that subsequently is separated into cleaned wastewater and thickened activated sludge. To meet the various effluent discharge standards, the secondary effluent is further treated to tertiary level. The tertiary treatment processes include coagulation/flocculation, sedimentation and sand filtration (which is a dead-end filtration).

The secondary sedimentation tank is the determining or limiting factor for the operation of the activated sludge process because the settling processes in secondary treatment mainly depend on the biomass concentration in the aeration tank. To achieve a safe separation of activated sludge and cleaned wastewater, the average biomass concentration in the activated sludge should be 3,000 to 4,000 mg/l and should not exceed 5,000 mg/l (Engelhardt et al., 1988). Not because these are optimal but because this is the maximum that can be achieved using gravity for solid/liquid separation in the secondary sedimentation tank.

2.6.4 Combined Activated Sludge Crossflow Membrane System

In this system, as shown in Figure 2.4, the membrane is kept outside the aeration tank. Washington et al. (1969) and Hardt et al. (1970) investigated the use of UF for the control of concentrated activated sludge in a lab scale size. MLVSS was recycled continuously for 8 days. The reactor was loaded with 0.2 kg COD/kg MLVSS. day of synthetic based glucose sewage. The COD removal obtained was more than 98%, and the average flux rate was 8 l /m^2 hr. A net decrease in concentration as well as the respiratory activity (as measured by the oxygen utilization rate) of the biological culture was observed as the end products accumulated. The low process loading and possible destruction of cells by the recirculation system could be a cause of death and subsequent lysis of inactive cells. Their results were inconclusive, since death and subsequent lysis of inactive cells could account for the loss in volatile suspended solids. However they concluded that ultrafiltration is highly effective means of separating microbial solids from the final effluent.

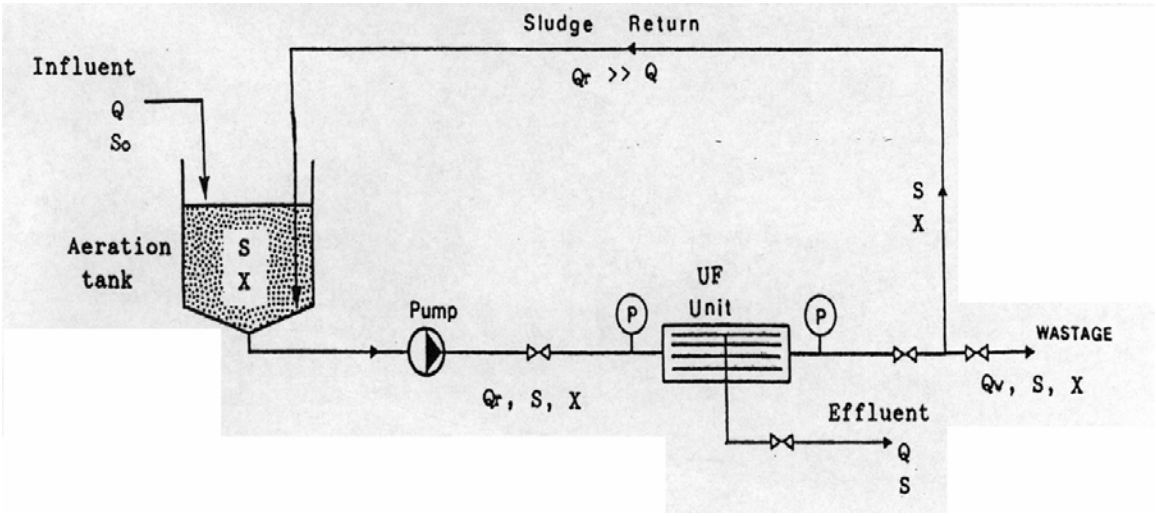

Figure 2.4: Schematic Diagram of Combined AS-UF system (El-Kebir, 1991)

Bemberis et al (1971) and Stavenger (1971) reported an AS/UF installation at Pikes Peak in Colorado. The installation was intended to be used for treating 1500 gpd of tourist effluent. During the trial, several advantages were reported:

- High effluent quality (BOD <1 mg/l, and SS beyond detection limits) was obtained
- Sludge waste was almost eliminated
- Less attention and skilled operators required, since the system can operate with wide a range of solids content.

El-Kebir (1991) studied the biokinetic coefficients of an activated sludge-ultrafiltration system. He made the following assumptions in the estimation of biokinetic coefficients:

i. The reactor is completely mixed
ii. The volume of the reactor is constant
iii. Complete rejection of MLSS by the membrane
iv. The substrate is not rejected by the membrane

Table 2.6 shows the Monod kinetic of different MLSS concentrations.

Table 2.6: Monod Kinetic Co-efficients at different MLSS concentrations

(El-Kebir,1991)

MLSS, mg/l	Y	k_d, (day^{-1})	μ_m, (day^{-1})	Ks(mgCOD/l)
10000	0.48	0.05	5.9	3720
5000	0.60	0.08	5.6	395
3000	0.47	0.16	8.10	250

Suidan M. T. (2000) studied performance of combined activated sludge crossflow microfiltration under various SRTs. He used synthetic feed to simulate moderate organic strength municipal wastewater and ceramic (alumina / titania-zirconia) membrane. The membrane was regenerated on weekly basis. He found biomass production rates increasing steadily with decreasing SRT and Higher fraction of the biomass were viable at lower SRTs. Effluent quality was maintained at sludge ages from 30 days to 5 days.

The effect of operating parameters such as transmembrane pressure and feed velocity on filtration flux was studied by Riesmeier et al. (1989), Matsumoto et al (1990), Al-Malack and Anderson (1997), and Miaomiao and Lianfa (2000). Shear breakage of microbial cells due to crossflow microfiltration was studied by Yasutoshi et al. (1994). The cell breakage due to shear stress reduced the filtration flux because of the increase of the hydraulic resistance of the particle packed layer, which was formed on the membrane during filtration.

2.6.5 Submerged Membrane Activated Sludge Process

A submerged membrane bioreactor, in which membranes are directly submerged into an aeration tank, (Figure 2.5), was first developed by Yamamoto et al. (1989). Early interest in membrane use for biomass retention in an activated sludge process had started mainly from Japan and also used in Europe and Canada. Membranes obviate the need for both primary and final sedimentation tanks thereby resulting in a considerable space saving. Additionally, high levels of MLSS were reported to effectively achieve nitrification and denitrification without the need for extended aeration (Magara et al. 1992). They described a pilot trial for collective night soil biological denitrification using a tubular

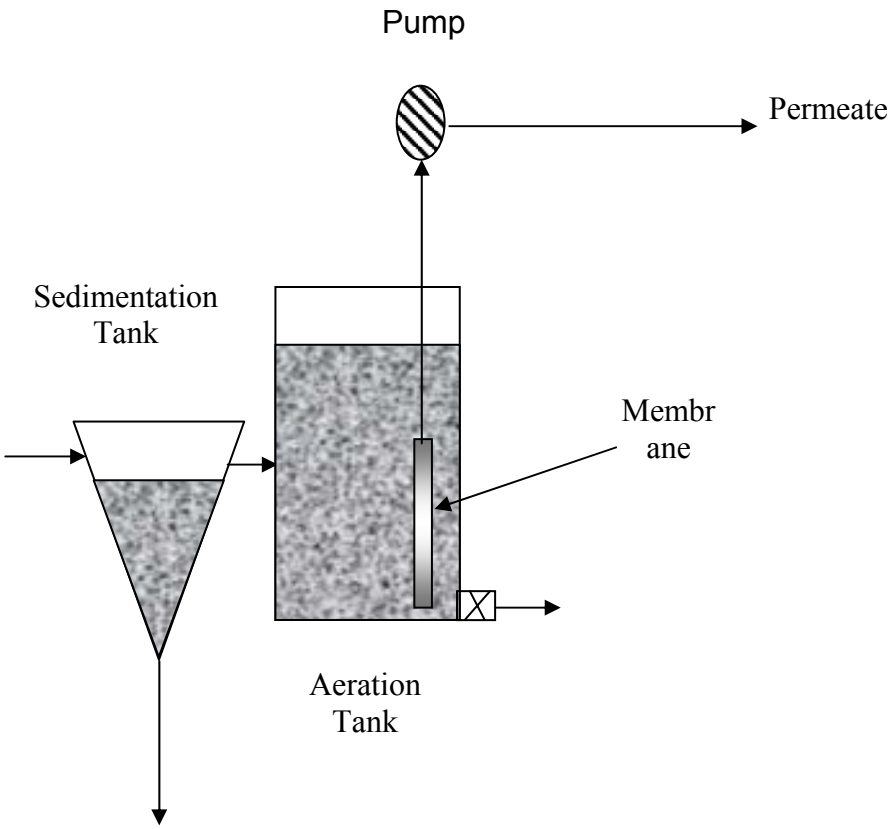

Figure 2.5: Submerged Membrane Activated Sludge Process

membrane in an activated sludge process. These authors showed that the investment and operational cost of the membrane process treating 100 m^3/d collective night soil was less than or equal to a conventional system but required less land and supervision. Their bioreactor operated with a feed BOD of 7800 mg/l, MLSS of 18000 mg/l and HRT of 4.7 days.

Tubular membrane systems are quite energy intensive and while they are favorable for highly concentrated feed such as night soil, their economics are unlikely to be attractive for domestic sewage. For this reason, Chiemchaisri et al. (1992) investigated the use of the hollow fiber (HF) membranes (.03 and 0.1 μm) in an activated sludge process for domestic sewage. In this study HF bundle was immersed in the bioreactor, permeate was withdrawn by the application of a vacuum and sludge build up around the fibers was minimized by agitation of the liquid. It was reported up to 90 % of nitrogen removal and a mean permeate COD of 20.8 mg/l.

Davies et al. (1998) submerged 0.4 μm polyolefin membrane complete with integrated air diffusers into the aeration tank. Operating with an average MLSS of 16,000 mg/l and 4.5 hours HRT, the bioreactor produced a very high quality effluent with typical values of 4 mg/l and 5 mg/l for BOD and NH_4 – N, respectively. With the availability of the energy efficient membrane modules for microfiltration, many researchers are working to realize new process engineering in municipal waste treatment. Engelhardt, et al. (1998), Ogoshi and Suzuki (2000), Gander et al (2000), Yasutoshi et al (1996) studied the integration of immersed membrane filtration into an activated sludge process in municipal/domestic wastewater treatment. They all reported COD & BOD removal efficiencies higher than 95 %. Sludge concentrations varied from 15 to 25 gMLSS/l.

Effect of particle size distributions of activated sludges on microfiltration flux for submerged membranes was studied by Yasutoshi et al (1997), using 20 types of sludges, such as intermittently aerated, continuously aerated activated sludges and anaerobically digested sludges. The flux varied from 0.5 to 1.7 m^3/m^2.day, under the same MLSS and filtration conditions. The variation of the flux was quantitatively interpreted by considering the concentration of particles with sizes from 8 to 15 μm, which cause the lowest lift velocity values. The assumption that the particles of this specific size range controlled the flux was rationalized by the filtration model proposed for the filtration of the mixture of differently sized particles.

The following three companies are marketing the submerged MBR configuration: Mitsubishi Rayon Corporation (Mitsubishi, Japan), Zenon Environmental Systems, Inc. (Zenon, Canada) and Kubota Corporation (Kubota, Japan). Full-scale MBR processes exist world-wide. To date, the largest U.S. MBR installation treating municipal water is in Arapaho County, CO, U.S.A. It is a 1-MGD capacity retrofitted sequencing batch reactor (Mourato and Frenkel, 1998).

Advantages of submerged membrane bioreactor

Submerged membrane bioreactor offers following significant potential benefits:

1) They have a smaller footprint than conventional processes. The systems can be operated at high mixed liquor suspended solids concentrations, and, in general, the high MLSS concentrations will result in lower effluent organic matter concentrations. Therefore, a smaller reactor can be used for a given level of treatment (or organic

matter decomposition). The smaller reactor and the absence of primary and secondary clarifiers result in a compact system (Cote et al. 1997).

2) Sludge wasting and handling requirements can be reduced significantly. Using membranes instead of settling tanks to clarify the reactor effluent enables MBR processes to operate at long sludge ages. The frequency of sludge wasting, removal of material from the reactor for disposal controls the sludge age. In conventional activated sludge systems, the operating sludge age impacts the settling characteristic of the microbial population that develops in the biological reactor. The operating sludge age is often limited by the settling performance of sludge in the final clarifiers. Changes in the microbial population such as the development of pin floc or filamentous flock have little impact on the effluent quality (Zhang et al. 1996). Therefore, sludge wasting, and solids handling operations that result, can be performed as a batch operation after relatively long intervals. Several studies have shown that membrane bioreactors produce less sludge than conventional activated sludge processes (Krauth and Stabb, 1994; Ishida et.al., 1993; Chaize and Huyard, 1991). Cote et.al., 1997 have reported sludge wasting to be virtually eliminated in an MBR.

3) The system requires little operators assistance, and, in general, little knowledge of the microbiological aspects of the process for successful operation. In conventional treatment, sludge bulking and other changes in the activated sludge microbial populations can diminish the overall effluent quality. This possibility requires constant process supervision by qualified personnel. In the submerged membrane process, because the solids separation step is virtually independent of the microbial

population, the supervision and expertise required are reduced. This is a significant benefit at installations where personnel would not be available for frequent monitoring and adjustment. Also, only intermittent wasting of solids and the relatively small volumes of material generated further reduce the operator attention required.

4) In the separate-stage configuration, the circulation pumps that feed the membranes can consume considerable energy. This can lead to relatively high operating costs and may limit its application. Even in conventional activated sludge process, recirculation pumps contribute to considerable energy consumption. The submerged membrane bioreactors does not require a circulation pump, they can operate with significantly less energy (Zhang et al., 1996).
5) The increase in sludge retention time in the submerged membrane bioreactors is thought to be the most effective factor for the enhancement of nitrification rate. The increase in sludge retention time leads to retention of nitrifying bacteria, whose increase in much slower than the BOD oxidizing bacteria (Kishino et al., 1995).
6) Because of the small pore size of the membranes employed for solids separation, submerged membrane bioreactors able to remove a wide range of microorganisms. Many studies have demonstrated that a membrane bioreactor is extremely efficient in the removal of excreted bacteria (Krauth and Stabb, 1993). Chiemchaisri et al. (1992) observed 4 to 6 log removal of Qβ bacteriophages by the gel layer formed on the membrane and Winnen et al. (1996) demonstrated that MS-2 bacteriophages were retained in a membrane bioreactor. Similarly Yamamoto et al. (1994) observed removal rates of up to 2 log for T1 and Qβ bacteriophages in a membrane bioreactor.

Disadvantages of submerged membrane bioreactor

1) The membranes operating life has not been firmly established. Although the apparent widespread acceptance of the process in Japan indicates that membranes have some degree of long-term reliability, a literature review did not find quantification of the life span in terms of years or volume of waste treated.
2) The major disadvantage of submerged membrane bioreactors is membrane fouling or membrane clogging. Fouling results from the accumulation and attachment of particulate and dissolved material at the surface of the membrane, which causes a significant resistance to filtrations. Additionally, the presence of stringy material such as hair or rags would significantly reduce membrane operation. This could be a major consideration for application without fine screens or a high degree of primary treatment.
3) The submerged membrane configurations will periodically require some form of chemical membrane cleaning. This can be accomplished with chlorine solution, or sometimes by immersion in an acid bath. However, additional chemical storage and handling requirements may be undesirable at some facilities.
4) Though submerged membrane bioreactors operate under conditions that potentially require very little wasting of sludge, however, at some point, sludge wasting and disposal is inevitable. Relatively little information is available on the properties of the waste sludge generated by the submerged membrane bioreactors. Available information indicates that dewatering of sludge would be difficult because of the large fraction of smaller-size particles. It is expected that this could be corrected by chemical conditioning. Because the costs for conditioning and dewatering of large

quantities of submerged membrane bioreactor sludge may be significant, sludge handling issues at larger facilities should be carefully considered.

5) Though submerged membrane bioreactors have small footprints, these systems require module removal from the aeration tank's top. Therefore, a significant amount of clearance is required.

In the submerged membrane activated sludge process, membranes are used extensively, at low pressure and below critical flux, where fouling is minimal (Howell, 1995). This ensures simple, reliable and low-cost operation. Aeration in the reactor is the important factor governing the filtration conditions. The cake removing efficiency of the uplifting air flow is affected by the turbulence of the flow. An increase in the air flow rate partly stimulated the cake-removing efficiency, but there is a critical value beyond which any increase in the air-flow rate would virtually has any effect on the cake-removing efficiency (Ueda et al., 1997).

From the above literature review, it can be concluded that membrane bioreactors are rapidly gaining inroads into municipal wastewater treatment. Hybrid system of membrane technology and activated sludge process has numerous benefits over conventional wastewater treatment processes. More research is needed in this area of study. Behavior of microorganisms in the SM-AS process needs to be understood so that these types of treatment plants could be designed to meet required effluent standards. The current literature lacks in this regard. It is hoped that the present study would bridge this gap.

CHAPTER 3

RESEARCH OBJECTIVES

Chapter 2, the literature review, has covered the basic principles of activated sludge process in general and membrane bioreactors in particular. This has shown that extensive research has been carried out in the areas of activated sludge process modifications. The previous chapter also reviewed the status of the current research in the field of application of membrane filtration to activated sludge process. Though much work has been done in the case of membrane filtration coupled with ASP, where the membrane is kept out side the aeration tank, still much work needs to be done in the case where the membrane is kept inside the aeration tank (submerged membrane activated sludge process). The literature lacks information, especially regarding biokinetic coefficients in a submerged membrane activated sludge process.

Submerged membrane separation utilizes high biomass. These cause considerable stress to the microorganisms involved. This stress might affect them by lowering the microorganism viability or altering their activities in substrate removal. Viability is defined as the ability of cells to grow and reproduce. At high biomass concentration, conditions for growth and metabolism are less favorable due to hindered access to

nutrients, space limitations, and accumulation of the end products. These factors might contribute to the lowering of the biomass activities, and may also lead to compositions different from those usually encountered.

Submerged membrane activated sludge process started developing, as a new process since early nineties. There is still a lack of understanding of the interaction between the biological and filtration units. Most of the research carried so far was focused on operating pilot plants and studying the removal efficiencies, without considering the biokinetic coefficients. To understand this new process, thorough investigations need to be carried out to find the behavior of the process in general and biomass in particular at different MLSS concentrations and under different organic loading rates.

Based on the above discussion, the detailed objectives of this research are:

i) Assess the treatment efficiency of the submerged membrane activated sludge process and the performance of microfiltration membrane under different operating conditions (such as MLSS, Organic loading rates, Sludge retention time).

ii) Study the biokinetic coefficients and examine the influence of biomass concentration on the biokinetic coefficients

iii) Verify the mathematical model for the combined submerged membrane-activated sludge process (SM-ASP)

iv) Assess the performance of the submerged membrane activated sludge process under shock loading (high influent substrate concentration) conditions

v) Assess the ability of the system to withstand organic (phenol) and inorganic (Chromium) toxic elements.

It is believed that, this study will yield valuable information and provide more knowledge about the application of submerged membrane in solid-liquid separation, in general, and in the field of wastewater treatment, in particular. Also, with the kinetic data, it would be possible to design these types of unit processes, efficiently.

CHAPTER 4

MATERIALS AND METHODS

4.1 MEMBRANE FILTRATION UNIT

The membrane used throughout the experimental run was a polyester woven fabric of tubular configuration, filled with inert glass beads to prevent collapse of the walls when suction is applied. After filling with the glass beads, the tubular membrane inner diameter was 12.5 mm with an effective length of 10 cm. (Plate 4.1). Characteristics of the membrane fabric used are shown in Table 4.1. One end of the tube was closed and other end fitted with poly vinyl chloride stud, so that it can be easily connected to the pump with tubing. Twelve such membrane units were fitted to the plexi glass stand (Plate 4.2). Six cylindrical stone air diffusers, evenly spaced, were also fixed to this plexi glass stand. These diffusers prevent the clogging of the membrane with activated sludge also it serves the purpose of mixing the reactor contents and maintaining aerobic conditions in the reactor.

4.2 DESCRIPTION OF THE CONTINUOUS REACTOR

The continuous flow reactor used in this study is shown in Plate 4.3. Figure 4.1 shows a flow chart of the process. It consisted of following components:

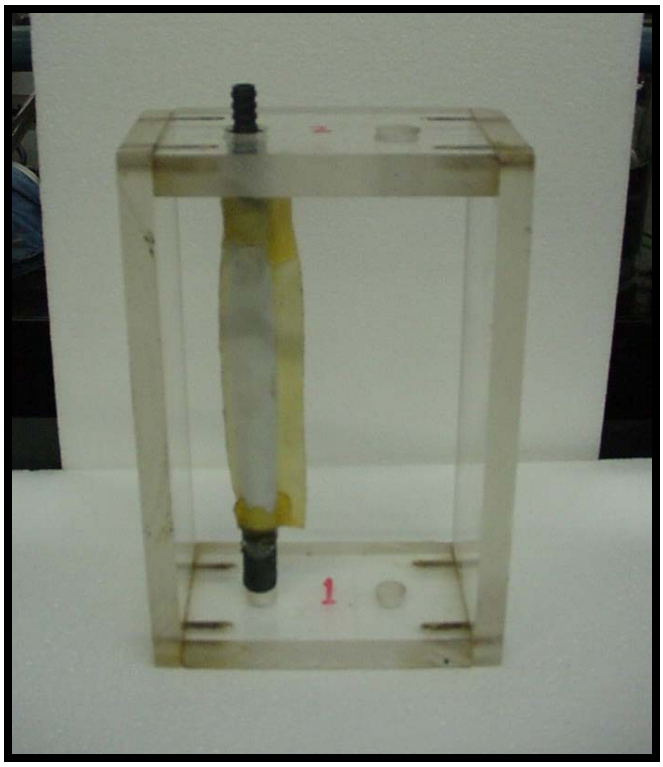

Plate 4.1: Polyester Woven Microfiltration Membrane

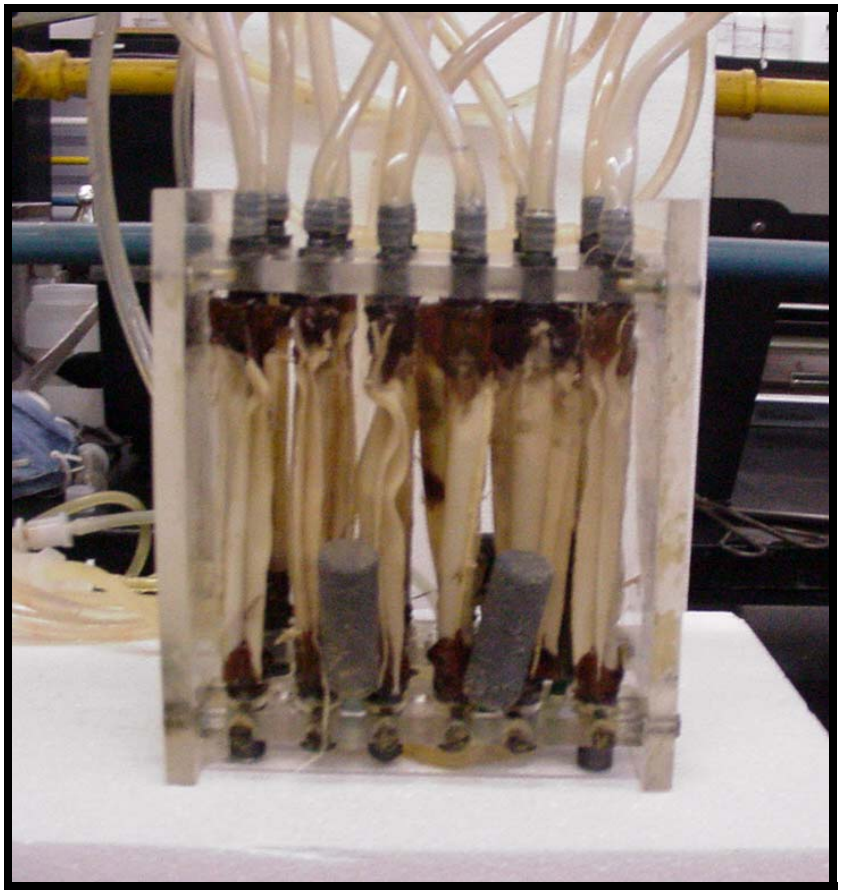

Plate 4.2: Membrane Filtration Unit

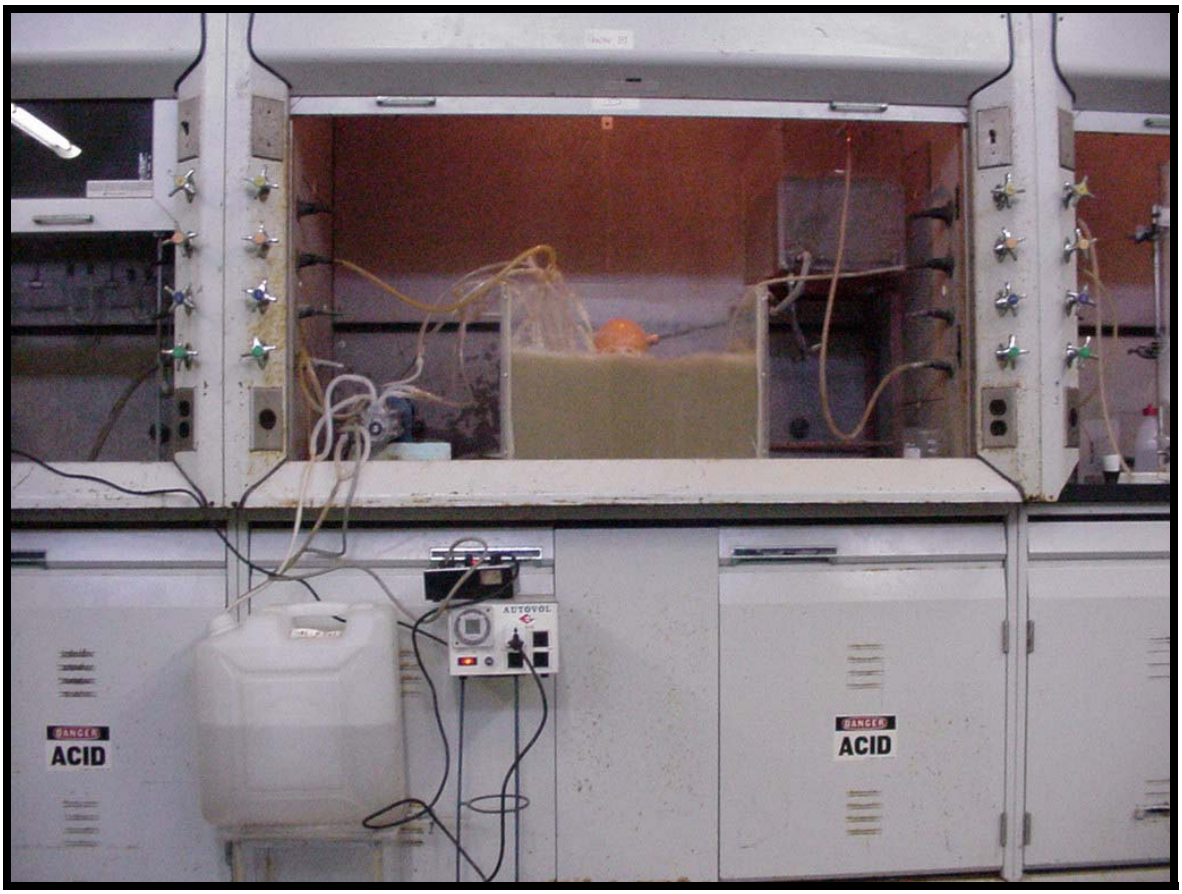

Plate 4.3: SM-AS System Setup

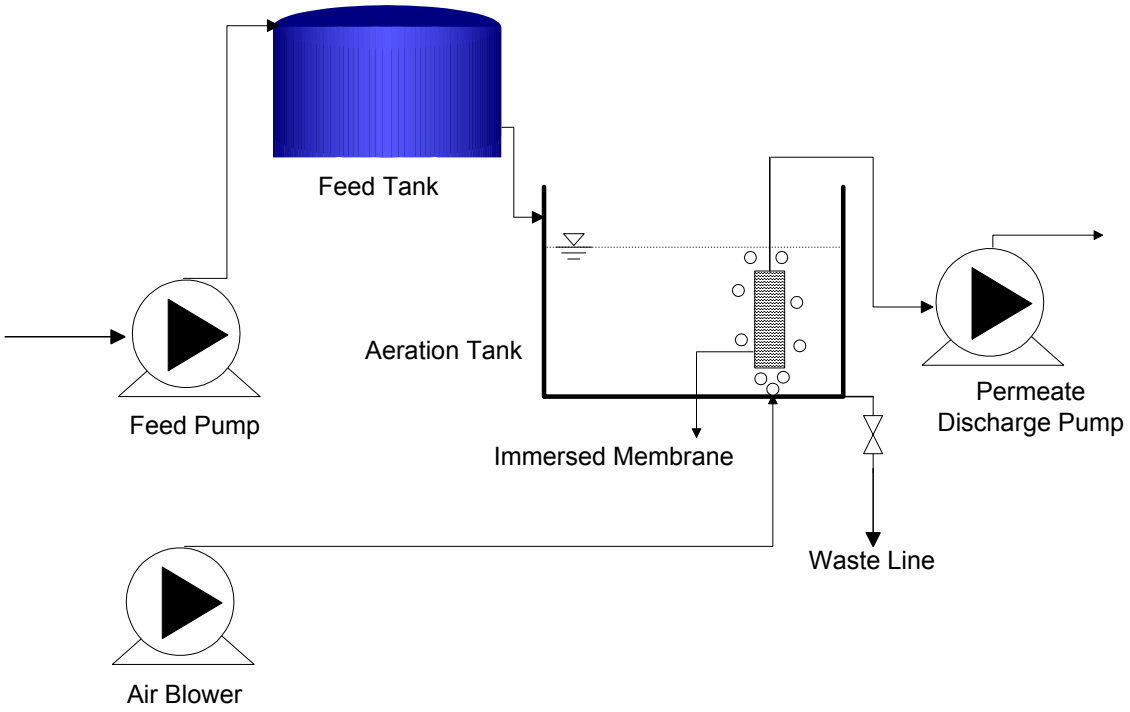

Figure 4.1: Experimental Flow Diagram

Table 4.1 Characteristics of the Polyester Woven Fabric

Configuration	Tubular
Material	Polyester
Pore-Size	20 to 40 μm
Diameter	1.27 cm (0.5 inch)
Length	12 × 10 cm
Cross – Sectional Area	1.27 cm^2
Total Surface Area	0.0479 m^2
Wall Thickness	0.34 – 0.375 mm
Hydraulic Resistance	2.7×10^6 l/m
pH Resistance Range	2 – 14

i. Feed Tank: This was a graduated, rectangular plexi glass tank of dimensions 41cm × 26cm × 30cm. Over flow outlet was positioned such that the capacity of the feed tank would be 28 liters. Concentrated synthetic wastewater was diluted with tap water as per influent concentration requirements in this feed tank.

ii. Aeration Tank: This was a rectangular tank with dimensions of 52cm × 21cm × 35cm. Wastewater was fed to this tank from the feed tank through a float to control the level of the wastewater inside the tank to 20 liters. The tank had an overflow arrangement and a waste drain.

iii. Diffusers: In total, eight cylindrical stone diffusers were used to keep the reactor contents under aerobic conditions. Six diffusers were used with membrane holding stand kept at the far end of the reactor and the remaining two diffusers were placed at the end near to the feed tank. These diffusers were connected to the air-injection line through Tygon tubes.

iv. Suction Pump: A Cole-Palmer peristaltic pump of variable speed was used at (seventy minutes “on” and five minutes “off”) intermittent operation to extract the permeate through membranes. Masterflex tubes were used to connect the pump to the microfiltration membranes.

v. Permeate Tank: A polyethylene container was used to collect the permeate. The container was graduated on the outside in order to facilitate the measurement of the permeate volume.

4.3 INFLUENT SUBSTRATE

A synthetic substrate consisting of Glucose, peptone and yeast extract as organic source was used as substrate throughout the research period to ensure a consistent quality of influent to the membrane bioreactor. The synthetic wastewater used in this study simulates municipal wastewater (Chang and Lee, 1998). It has been designed to provide all the inorganics and micronutrients, as well as nitrogen, phosphorous for the development of the biomass. The detailed composition of the synthetic substrate is shown in Table 4.2. Concentrated feed (100,000 mg/l COD) solution was prepared and stored in the refrigerator at 4 ^{0}C for a maximum period of seven days. Influent feed concentration of desired strength in terms of COD was then prepared by diluting the concentrated feed with tap water.

The influent substrate concentration varied from 500 mg/l COD to 4,000 mg/l COD for the biokinetic studies. Table 4.3 shows the general characteristics of the synthetic substrate. The influent was continuously supplied to the reactor in order to match with the permeate flow rate by keeping the water level constant in the reactor using a mechanical float. The seeding microorganisms were obtained from the return sludge at Saudi Aramco wastewater treatment plant, Dhahran.

4.4 MEMBRANE CLEANING

Membrane fabric cleaning was achieved by combination of various methods including intermittent pumping of the suction pump, continuous air scouring by air from stone aerators in the reactor, back washing with water, back washing with air and mechanical cleaning of the membrane fabric with a brush.

Table 4.2 Composition of the Synthetic Substrate

Component	**Contents in the Stock Solution**	**Contents in the Typical Feed Solution**
Glucose, $C_6H_{12}O_6$	40, 000	200
Peptone	40, 000	200
Yeast extract	4, 000	20
$(NH_4)_2\ SO_4$	32, 000	160
$KH_2\ PO_4$	6, 400	32
$MgSO_4.\ 7H_2O$	8, 000	40
$MnSO_4.\ 6H_2O$	720	3.6
$FeCl_3.\ 6H_2O$	40	0.2
$CaCl_2.\ 2H_2O$	800	4
COD (mg/l)	**100, 000**	**500**

Table 4.3 General Characteristic of Synthetic Wastewater

Constituent	**Concentration**
Ph	7.0
Suspended Solids	< 1 mg/l
BOD	380 – 3000 mg/l
COD	500 – 4000 mg/l
TOC	131.5 – 1000 mg/l

4.5 ANALYTICAL PROCEDURES

For the continuous reactor experiments, the sampling from the reactor and permeate were carried out periodically and analyzed for the following Physio-chemical parameters, by the methods described in the Standard Methods for the examination of wastewater (APHA.AWWA.WEF, 1998).

4.5.1 Turbidity

Turbidity is the interference of light passage through a sample by scattering and adsorption and is caused by colloidal particles such as microorganisms and insoluble organics. An Orbeco-Hellige digital direct reading turbidity meter was used to determine the turbidity of permeates.

4.5.2 Temperature and pH

The temperature was not controlled, however the temperature in the reactor was monitored regularly through out the duration of the experiments. Values of pH were measured using a pH meter. Phosphate buffer solution was used to maintain pH at 7.

4.5.3 Suspended Solids

Suspended solids removal is considered to be an important factor when assessing the operational performance of microfiltration processes in wastewater treatment. Also, concentration of biomass in the aeration tank is represented by mixed liquor suspended solids (MLSS) and mixed liquor volatile suspended solids (MLVSS).

Samples were tested for suspended solids by filtering a measured volume of the sample through filter paper of 0.45 μm pore size and oven drying for at least one hour at 105°C and cooling in a desiccator to a constant weight. Volatile suspended solids were determined by igniting for 15 to 20 minutes, the filtered solids, at 500°C ± 50°C in an electric muffle furnace.

4.5.4 Dissolved Oxygen

Dissolved oxygen is required for the respiration of aerobic microorganisms as well as other aerobic life forms. A minimum dissolved oxygen concentration of 1 to 2 mg/l throughout the aeration tank is to be maintained so as to prevent anaerobic conditions to prevail in the reactor. Dissolved oxygen probe manufactured by Hanna Incorporations was used for measuring the DO in the aeration tank.

4.5.5 Chemical Oxygen Demand

Chemical Oxygen Demand (COD) is defined as the amount of a specified oxidant that reacts with the sample under controlled conditions. The quantity of oxidant consumed is expressed in terms of its oxygen equivalence. Because of its unique chemical properties, dichromate ion ($Cr_2O_7^{2-}$) is the specified oxidant used. COD of the influent, reactor contents and permeate were analyzed periodically using closed reflux, tritrimetric method.

4.5.6 Biochemical Oxygen Demand

Biochemical Oxygen Demand (BOD) is a measure of the quantity of the oxygen consumed by bacteria when stabilizing degradable organic matter under aerobic conditions according to following equation:

$$\text{Organic Matter} + \text{Dissolved Oxygen} \xrightarrow{\text{Bacteria}} CO_2 + \text{Biological Growth} \qquad (4.1)$$

The BOD_5 tests were carried out on permeate, reactor and feed samples according to Standard Methods (APHA.AWWA.WEF, 1998). 300 ml BOD bottles were used. The samples were suitably diluted. Dissolved oxygen probe manufactured by Hanna Incorporation was used for initial and final dissolved oxygen measurements.

4.5.7 Total Organic Carbon

The Total Organic Carbon (TOC) test measures the organically bound carbon in a wastewater sample. The test is performed by injecting a known quantity of sample into a high temperature furnace or into a chemically oxidizing environment. The organic carbon is oxidized to carbon dioxide in the presence of a catalyst. The carbon dioxide that is produced is quantitatively measured by means of an infrared analyzer (Metcalf & Eddy, 1991). The TOC was measured using TOC analyzer.

4.5.8 Phenol

The phenolic solution used in the toxic organic loading experiments was buffered by adding 1.3 g/l of KH_2PO_4 and adjusted to pH of 7.0. Phenol concentrations were determined using UV-1601 PC UV-Visible spectrophotometer manufactured by

Shimadzu. The permeate samples were filtered using filter paper number 42 and pH was measured and adjusted (Essa, 1993). The wavelength used was 269 nm.

4.5.9 Chromium

Chromium metal was used for the toxic inorganic loading experiments. Chromium was determined by atomic absorption spectrophotometer at 357.9 nm with air-acetylene flame. On collection, the samples were acidified to pH<2 with concentrated nitric acid.

4.6 MICROBIOLOGICAL PARAMETERS

4.6.1 Heterotrophic Plate Count

The Heterotrophic Plate Count (HPC), formerly known as the standard plate count, is a procedure for estimating the number of live heterotrophic bacteria in wastewater samples. Plate count agar (tryptone glucose yeast agar) was used as media. The 15 ml of plate count agar was autoclaved at 1.05 kg/cms and 121°C for 20 minutes and then distributed in petri dishes along with the diluted sample after thorough mixing. The samples were distributed in three to seven dilutions and in triplicates. The petri dishes were then incubated at 35°C for 48 hours (APHA.AWWA.WEF, 1998). After the incubation period, the petri dishes were examined under a colony counter and the colonies were enumerated. The dilutions that yielded 30 to 300 colony-forming units (CFU) only were selected.

4.6.2 Most Probable Number

The Most Probable Number (MPN) is an index of the number of coliform bacteria that, more probably than any other number, would give the results shown by the laboratory examination. Total coliform, fecal coliform and fecal streptococci in the reactor and permeate, were analyzed using multiple fermentation tubes, according to Standard methods (APHA.AWWA.WEF, 1998).

4.6.3 Bacterial Species Identification

The bacterial colonies from the heterotrophic plate counts were further analyzed at microbiology lab at King Faisal University, Dammam, to identify the species of the bacteria. The identification was carried out using API 20E kit, (Analytical Profile Index) Analy Lab products. The API 20E is a standardized identification system and consists of 20 micro tubes containing appropriate dehydrated media. These micro tubes were inoculated with the isolated bacteria for 48 hrs at 20°C. During the incubation, metabolism produces color changes. The reactions are read according to the interpretation table provided, either positive or negative, which in tern generates a nine-digit number. This number then provides the characteristic of each species.

4.7 EXPERIMENTAL PROCEDURE LAYOUT

Initial seeding of the bioreactor was accomplished by charging the bio-reactor with 20 lit of the returned activated sludge from the Saudi Aramco wastewater treatment plant, Dhahran. The MLSS of this mixed liquor was 3000 mg/l. The first four weeks of the operation were allowed for the biomass to acclimatize to the synthetic substrate. The

pumping was carried out intermittently with the schedule as shown in Table 4.4. Intermittent suction method was applied because intermittent suction showed higher performance than continuous suction with regard to the maintenance of stable flux (Yamamoto et al., 1989). GRASSILIN auto timer achieved the intermittent schedule. The experimental runs were carried out for duration over a year, with continuous monitoring to establish biokinetic coefficients and to assess the ability and suitability of the membrane bioreactor to provide the required COD removal and to absorb the shock loadings.

The experimental investigation in this study consists of two phases. In the first phase, the biokinetic co-efficients were determined by operating the system at various sludge retention times (SRT) and by allowing (at each adopted stage of SRT) a steady state condition to prevail. At the beginning of the study, an MLSS concentration of 3000 mg/l was attained and maintained under steady state conditions. A steady state condition is achieved when fairly constant biomass growth & filtrate COD are obtained. Sludge was wasted daily to maintain steady state conditions. Accurate measurement of the biomass and effluent substrate concentrations were recorded.

Then, by increasing the Organic loading rate (OLR, kg COD/ kg MLSS day) and controlling the SRT, a second steady state condition for same MLSS concentration was achieved and biomass and effluent substrate concentration were recorded. Then, third steady state condition was established by further increasing the OLR. The kinetic coefficients were determined by plotting these parameters at steady state conditions.

Table 4.4 Pumping Schedule

ON TIME	OFF TIME
08.00 AM	09.10 AM
09.15 AM	10.25 AM
10.30 AM	11.40 AM
11.45 AM	12.55 PM
01.00 PM	02.10 PM
02.15 PM	03.25 PM
03.30 PM	04.40 PM
04.45 PM	05.55 PM
06.00 PM	07.10 PM
07.15 PM	08.25 PM
08.30 PM	09.40 PM
09.45 PM	10.55 PM
11.00 PM	12.10 AM
12.15 AM	01.25 AM
01.30 AM	02.40 AM
02.45 AM	03.55 AM
04.00 AM	05.10 AM
05.15 AM	06.25 AM
06.30 AM	07.40 AM

Next, the biomass concentration was increased to 5,000 mg/l, 10,000 mg/l and 15,000 mg/l and similar analysis were carried out after attaining steady state conditions at each of the specified substrate concentrations. The organic loading rate ranged from 0.40 kg COD/ kg MLSS day to 5 kg COD/ kg MLSS day in the above mentioned experimental runs. Flux was also monitored to assess the performance of the submerged microfiltration membrane under different MLSS concentrations and different organic loading rates.

In the second phase, at the 15, 000 mg/l MLSS concentration, the reactor was subjected to a shock loading (a high influent substrate concentration) of 15, 000 mg/l COD and effluent quality was monitored to study the ability of the submerged membrane activated sludge process to with stand shock loadings.

After this, the reactor was fed with synthetic substrate contaminated with phenol. The effect of phenolic toxic loading on the performance of the submerged membrane activated sludge process was studied without acclimatization as well as with acclimatization. For the acclimatization experiments, the bacteria were acclimatized to the phenol for over a month and then effluent quality as well as was bacterial populations in the reactor was monitored to study the effect of organic toxic pollutant to the submerged membrane activated sludge process. Then the reactor was fed with synthetic substrate contaminated with chromium. The effluent quality as well as was bacterial populations in the reactor was monitored to study the effect of inorganic toxic pollutant to the submerged membrane activated sludge process.

CHAPTER 5

RESULTS AND DISCUSSION

5.1 INTRODUCTION

The experimental runs were started on September 15, 2001, and continued for a period over a year. The entire duration of experiments can be divided into following stages:

- Acclimatization and pumping schedule
- Membrane fouling control
- Biokinetic coefficients
- Microbial analysis
- Shock loading and Toxic loading (phenol and chromium)

5.2 ACCLIMATIZATION AND PUMPING SCHEDULE

As mentioned in the previous chapter, the return activated sludge was brought from Saudi Aramco wastewater treatment plant and acclimatized to the glucose-peptone based synthetic substrate for one month. The concentration of mixed liquor suspended solids was 3000 mg/l and was maintained the same for the entire acclimatization period by wasting the excess sludge on daily basis.

The membrane unit consisted of twelve polyester oven fabric filters filled with glass beads (Plate 4.2). The total surface area of filtration was 0.0479 m^2. A number of trail experiments were conducted to determine the intermittent pumping schedule. The criterion for selecting the pumping schedule was to maintain a cumulative average flux of 34 $l.m^{-2}.hr^{-1}$. This average flux, intern will result in hydraulic detention time (θ) of the aeration tank around 12 hours. Intermittent pumping would improve the flux and retard the rate of flux decline (Yoon et al., 2000 and Kishino et al., 1995).

After several trials, an intermittent pumping operation of seventy minutes "on" and five minutes "off" schedule were found to meet the desired average permeate flux criteria. The pumping schedule for a day is shown in Table 4.4. Figure 5.1 shows the variation of the flux in a day with the clean membrane and Figure 5.2 shows the variation of the effluent turbidity with the clean membrane in a day (Table D1). As evident from these Figures, the flux dropped from 745 $l.m^{-2}.hr^{-1}$ to 187 $l.m^{-2}.hr^{-1}$during the first on-off cycle. In the second on-off cycle, the flux changed from 238 $l.m^{-2}.hr^{-1}$ to 52 $l.m^{-2}.hr^{-1}$ and in subsequent cycles it varied from 56 $l.m^{-2}.hr^{-1}$ to 28 $l.m^{-2}.hr^{-1}$, with an average effluent flux of 34 $l.m^{-2}.hr^{-1}$. Correspondingly, the permeate turbidity dropped from 185 NTU to 2.8 NTU during the first on-off cycle. In the second on-off cycle, the effluent turbidity changed from 5.2 to 0.5 and in subsequent cycles it varied from 0.55 NTU to less than 0.01 NTU.

During the typical operation period, the variation of the flux and variation of the turbidity in a day are shown in Figure 5.3 and Figure 5.4 respectively (Table D2). The steady drop in the permeate flux was controlled and flux rate enhanced by the intermittent pumping mechanism. This can be clearly seen in the Figure 5.3, the peaks at the interval of seventy

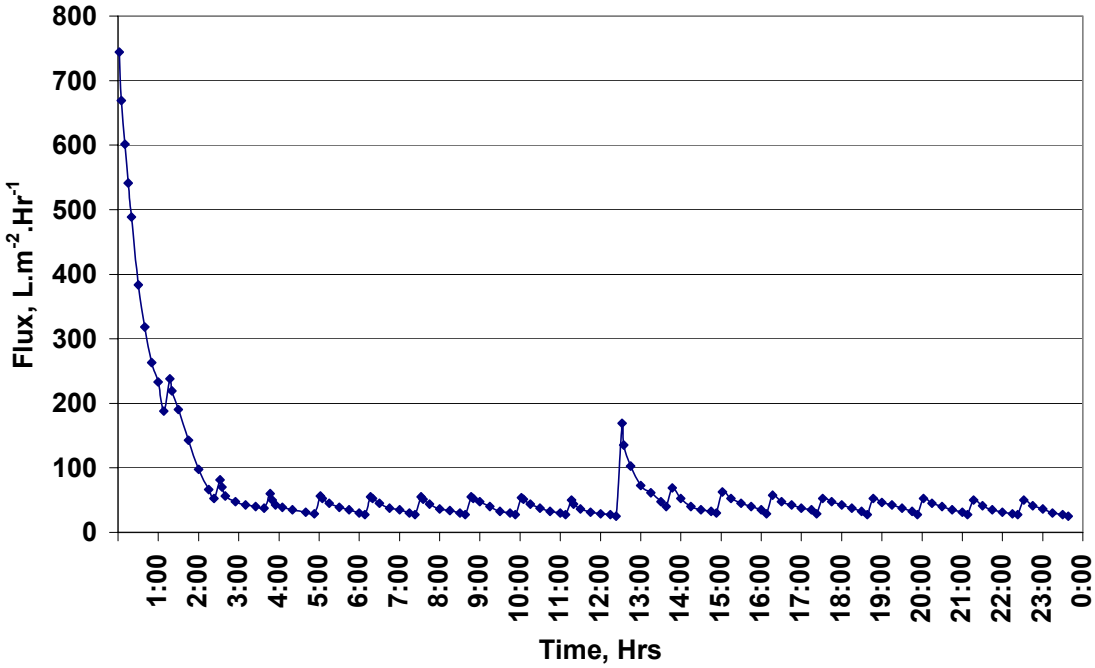

Figure 5.1: Variation of Flux in One Day, With the Clean Membrane

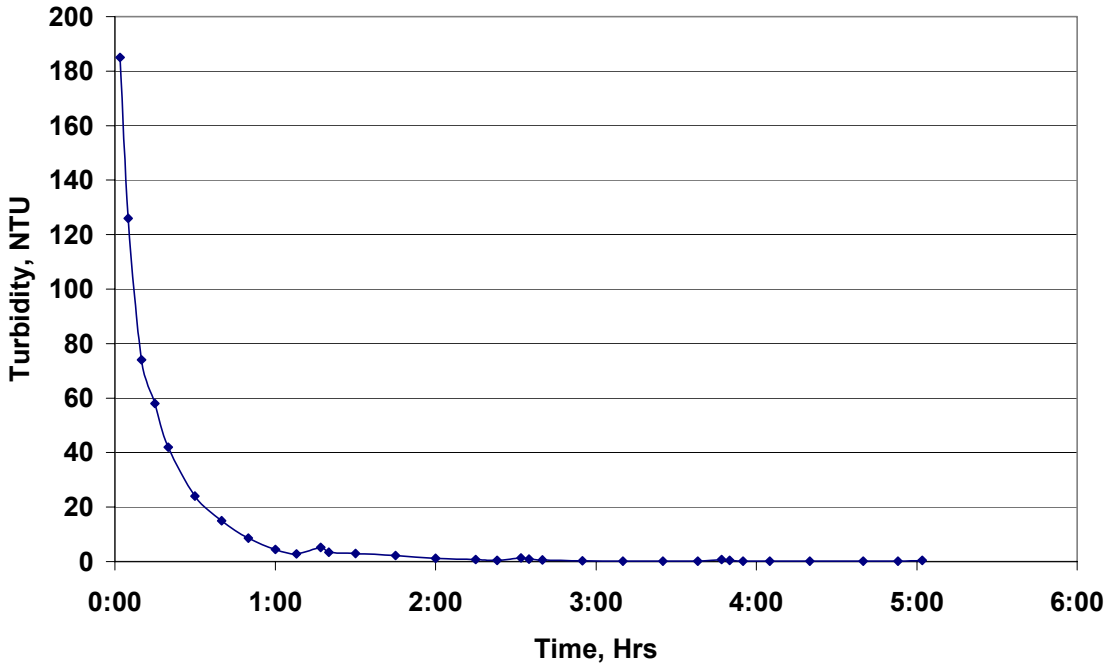

Figure 5.2: Variation of Effluent Turbidity with the Clean Membrane

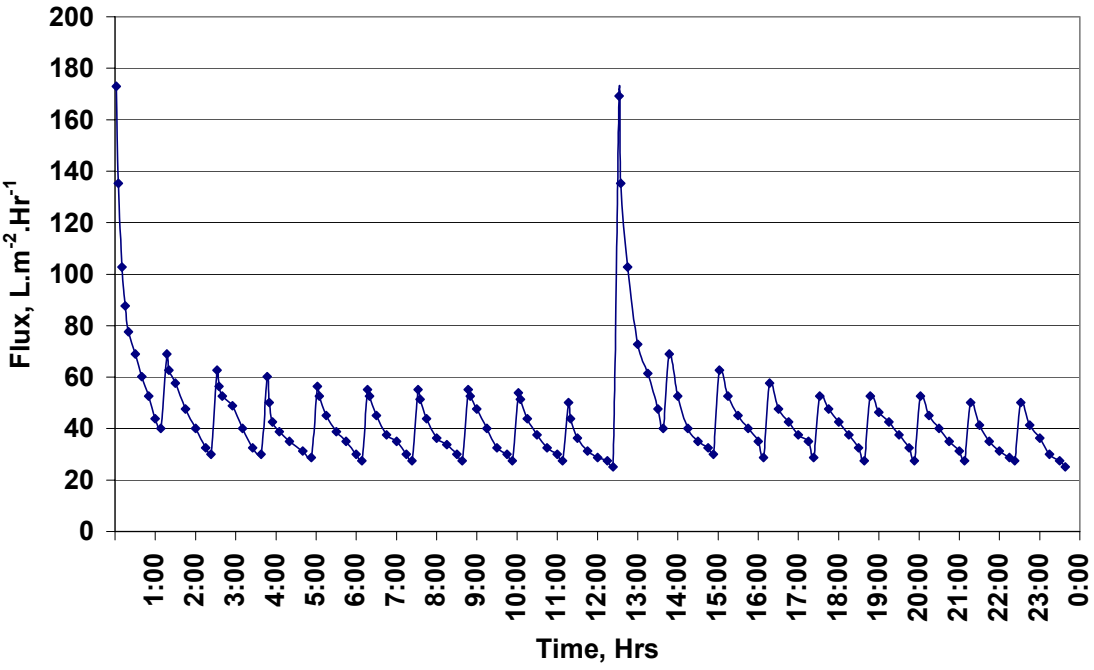

Figure 5.3: Typical Variation of Flux with Time, in One Day

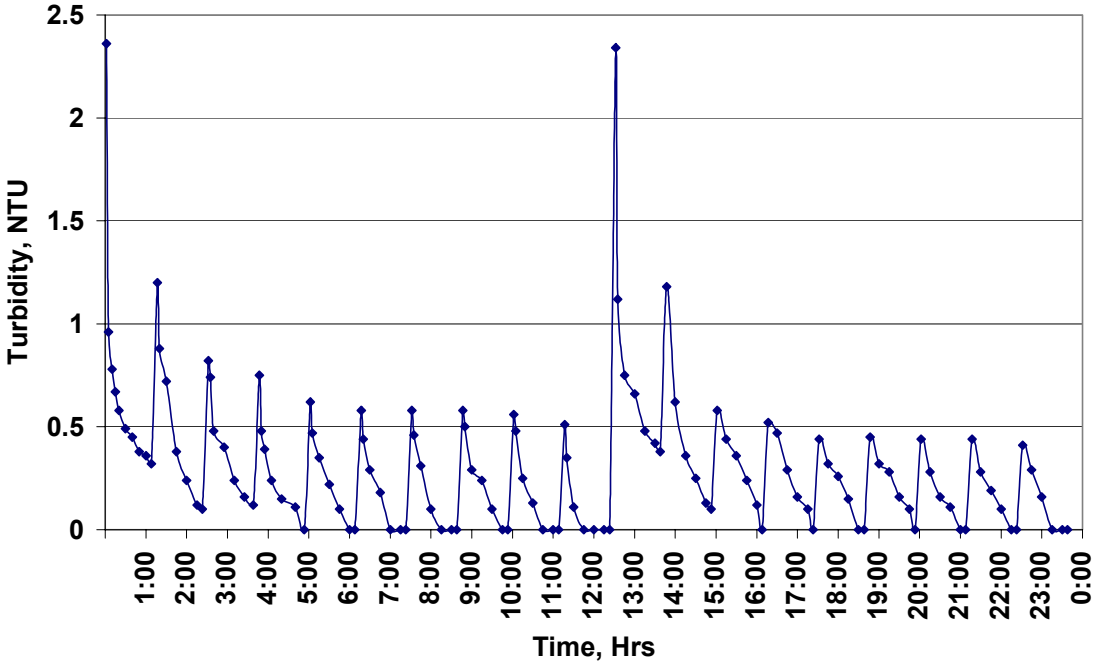

Figure 5.4: Typical Variation of Permeate Turbidity with Time

minutes indicate the starting of a new on-off cycle. It is significant to note that, though by intermittent pumping, the rate of permeate flux enhanced, but it did not significantly increase the effluent turbidity. In other words, the permeate quality was not significantly changed by the intermittent pumping schedule.

When the permeate flux dropped to lower than 25 $l.m^{-2}.hr^{-1}$, and an average flux of 34 $l.m^{-2}.hr^{-1}$ could not be maintained, then the pump was stopped and the membranes were back flushed with air for three minutes. This back washing with air was performed after every twelve hours of operation. The effect of back flushing the membrane with air can be clearly seen in Figure 5.3, the permeate flux improved to 169 $l.m^{-2}.hr^{-1}$ from 25 $l.m^{-2}.hr^{-1}$.

5.3 MEMBRANE FOULING CONTROL

The membrane fouling which occurred during the operation can be observed by monitoring the flux. As the suction is continued in the bioreactor, a cake layer accumulated on the membrane surface, causing decline in the flux. To observe the membrane fouling, the fouled membrane samples were taken out and were observed. As shown in plate 5.1, the foulants adhered on the outer surface of the membrane. On the outer surface of the membrane, there existed a sludge layer consisting of microorganisms. Beneath the sludge layer was a sticky gel layer, probably formed by the adsorption of soluble matter. To keep the permeate flux within the required limits; two kinds of counter measures for cake removal were undertaken. One is the continuous measure and other is the intermittent measure.

Plate 5.1 Plate Showing Fouled Membrane.

5.3.1 Continuous Measure

In the aeration tank, air was supplied by two stone diffusers, which are submerged at the near end of the aeration tank. The main purpose of these air diffusers was to provide oxygen to the aeration tank and to give turbulence for mixing the contents of the reactor. Another set of six stone diffusers were installed under the membranes because cake layer on membrane surface was to be removed by a sheering stress generated by the uplifting flow of bubbling air. The main purpose of these air diffusers was to give crossflow filtration effect. This arrangement also gave turbulence and oxygen to the reactor.

The turbulence existing in the aeration tank and the sweeping effect of the air bubbles were sufficient to give the conditions of subcritical operation due to low effluent permeate fluxes (about 35 $l.m^{-2}.hr^{-1}$) (Chiemchaisri et al., 1992 and Ueda et al., 1997). Thus with this continuous measure generally a low (25 $l.m^{-2}.hr^{-1}$ – 40 $l.m^{-2}.hr^{-1}$) but fairly constant flux was obtained (Figure 5.1 and Figure 5.3).

5.3.2 Intermittent Measure

Intermittent Suction:

Intermittent suction method was adopted as a measure of membrane fouling control. After operation of every seventy minutes, the suction was suspended for five minutes. However, aeration using stone diffusers beneath the membranes was continued. During the suction suspension, the cake layer might be removed successfully, because suspended solids in mixed liquor could not concentrate upon the membrane surface when suction was absent (Ueda et al., 1997).

Also, sudden off and on of suction pressure might produce mechanical shocks that cause loosening of the cake layer and subsequent easy removal by the sweeping action of the air bubbles beneath. The adoption of this "aeration without suction" measure was found to be very effective and convenient way of in-situ membrane washing without taking the membrane modules outside the reactor. With this technique alone, average flux of 34 $lm^{-2}hr^{-1}$ was achieved for almost twelve hours. The minor peaks after every seventy five minutes in Figure 5.1 and Figure 5.3 show increase in flux due to sudden start of suction after stopping the pump for five minutes.

An important point observed with intermittent pumping schedule is, though it improved flux, but it did not increase the turbidity in the effluent drastically. Even after the sudden "on" of suction, the effluent turbidity was less than 0.6 NTU, as shown in Figure 5.4. This shows that intermittent suction operation enhances the permeate flux without degrading the permeate quality. After twelve hours, flux dropped further and reached below the critical flux of 25 l $m^{-2}hr^{-1}$ and this "on and off" mechanism alone was not enough to bring the flux to the desired levels. So, to improve the flux further, back flushing with air was adopted.

Back Flushing:

After over twelve hours of operation, the fouling of the membrane was so severe that the intermittent suction operation alone was not sufficient to loosen and remove the cake layer formed around the membrane to give the desired flux. To solve this fouling problem, three in situ back flushing with air and water procedures were tested.

i. Back flushing with water alone: When the suction was stopped during the "on-off" cycle, the flow direction of the pump was reversed, and permeate collected in the permeate tank was back flushed for two minutes through the membranes. Then the pumping was stopped for the next three minutes and the pump direction was reversed. The effluent flux was monitored when the pump was on during the start of next "on-off" cycle. It was observed that during the back washing with water, the cake layer which was adhering to the outer walls of the membrane were loosened and were scoured by the bubbling air from below the membrane unit, thus improving the flux considerably from below 25 l $m^{-2}hr^{-1}$ to 150 l $m^{-2}hr^{-1}$.

ii. Back flushing with water and air: After another twelve hours of operation, when the flux dropped below the critical flux limit, the pump was stopped and flow direction reversed and membranes were back washed with permeate as explained previously, but this time for one minute only. After this, the pump was stopped and the tubes were disconnected from the pump and connected to air supply. The air was forced through the membranes for one minute and then, tubes were connected back to the pump. The effluent flux was monitored when the pump was "on" during the start of next "on-off" cycle. It was observed that the flux improved from below 25 l $m^{-2}hr^{-1}$ to 180 l $m^{-2}hr^{-1}$.

iii. Back flushing with air alone: After another twelve hours of operation, when the flux dropped below the critical flux limit, the pump was stopped and the tubes were disconnected from the pump and connected to air supply. The air was forced through the membranes for two minutes and then, tubes were connected back to the pump. The effluent flux was monitored when the pump was "on" during the

start of the next "on-off" cycle. It was observed that during the back flushing with the air, the cake layer which was adhering to the outer layer of the membrane was loosened and scoured off by the back flushing air as well as the air bubbles from beneath the membrane unit. The flux enhanced substantially from below 25 l $m^{-2}hr^{-1}$ to 169 l $m^{-2}hr^{1}$.

Figure 5.5 shows the improvement in the flux because of each of these back washing procedures. The variation of permeate turbidity as a result of these back washing techniques is shown in Figure 5.6. All the three back flushing techniques gave satisfactory improvement in the permeate flux and this flux was well maintained for another twelve hours by intermittent pumping operation. Also, from Figure 5.6, it is evident that the turbidity of the effluent has not increased significantly, and within one hour of operation the turbidity dropped to below 0.5 NTU (Table D3).

Although all the three back washing techniques gave satisfactory results, back flushing the membrane with permeate was discarded because, back washing with water means increase in the quantity of the wastewater to be treated. Also, it was observed that reversing of the pump flow, for back washing with the permeate, would result in faster damage to the pump's tubing resulting in increase in cost of replacement as well as loss of time that is needed to replace the pump's tubing. On the similar grounds, back washing with air and water procedure was also not adopted.

Finally, from the above experiments it was concluded that back flushing the membrane with air for two minutes after every twelve hours of pumping was chosen to be adopted for the entire duration of the experiments, as this method was easy for operation and also

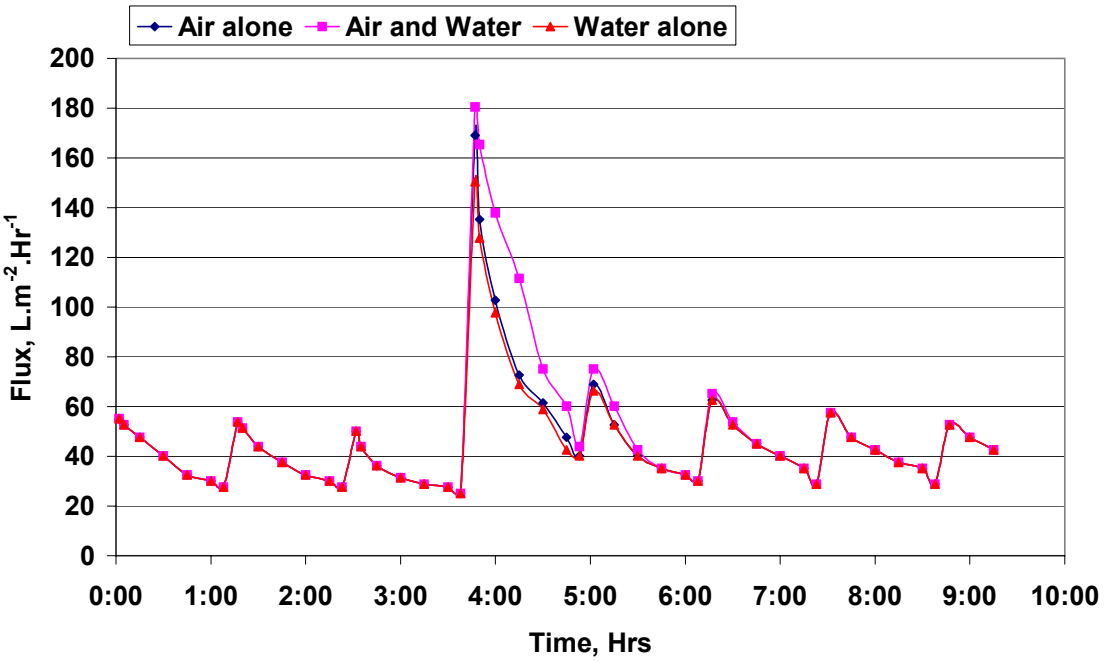

Figure 5.5: Variation of Flux Before and After Back Flushing

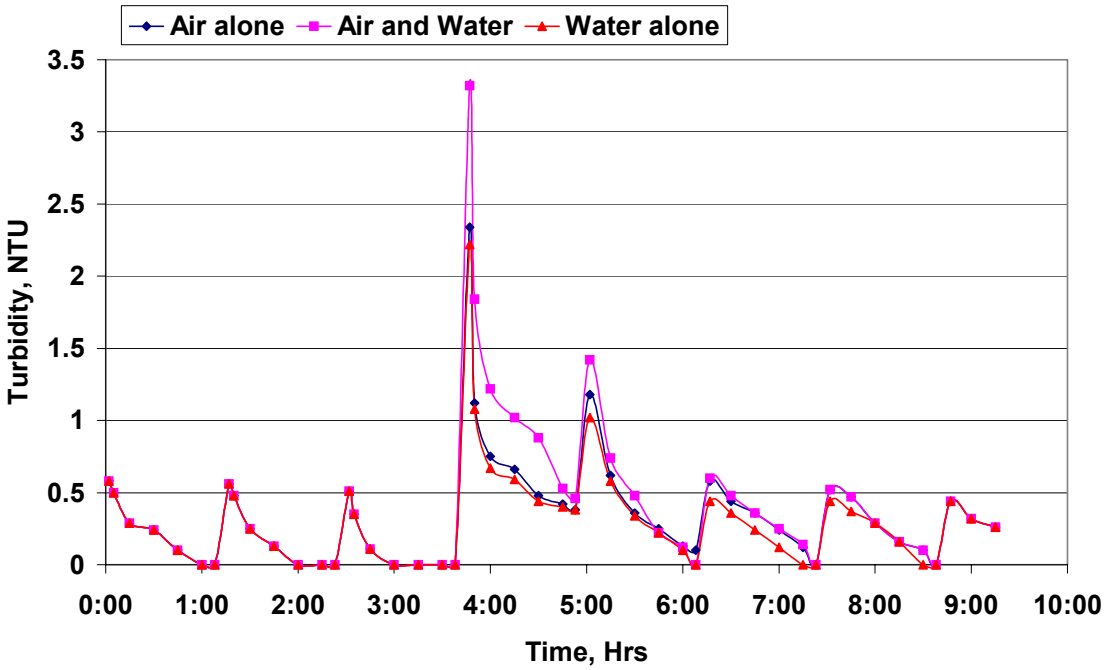

Figure 5.6: Variation of Turbidity Before and After Back Flushing

economical as compared with other two discussed methods. The back flushing technique not only improved the flux, also the permeate quality in terms of turbidity was satisfactory.

Brushing the membrane:

The combination of intermittent pumping and periodic back flushing with air was able to achieve the desired critical flux for a week. After seventh day, the critical flux could not be achieved by above mentioned membrane cleaning techniques. This could be probably due to sticky gel layer formed by the adsorption of soluble matter. Though the sludge layer that adhered on the outer surface of the membrane was partially cleaned by the intermittent suction and back flushing with air technique, the stubborn sticky layer that formed beneath the sludge layer was not dislodged by the above methods. To achieve the desired flux at this point, the membranes were brushed on the outer surface as mentioned below.

The pumping was stopped and the flow direction was reversed. The membrane filtration unit was taken out of the aeration tank and placed in a plastic tub. Ordinary tooth brush was used to clean the membranes. Plate 5.2 shows the brush at the time of cleaning. This practice employs the use of brush on the outside surface of the woven fabric and took place while back circulating the permeate water at low flow rate. This method of cleaning was adopted to ensure flushing out the sticky gel layer both on the inside surface of the fabric and within the weave. After cleaning was carried out, it was observed that the fabric has been reinstated to its previous condition that is, the flux obtained was same as the flux obtained initially with the clean membrane.

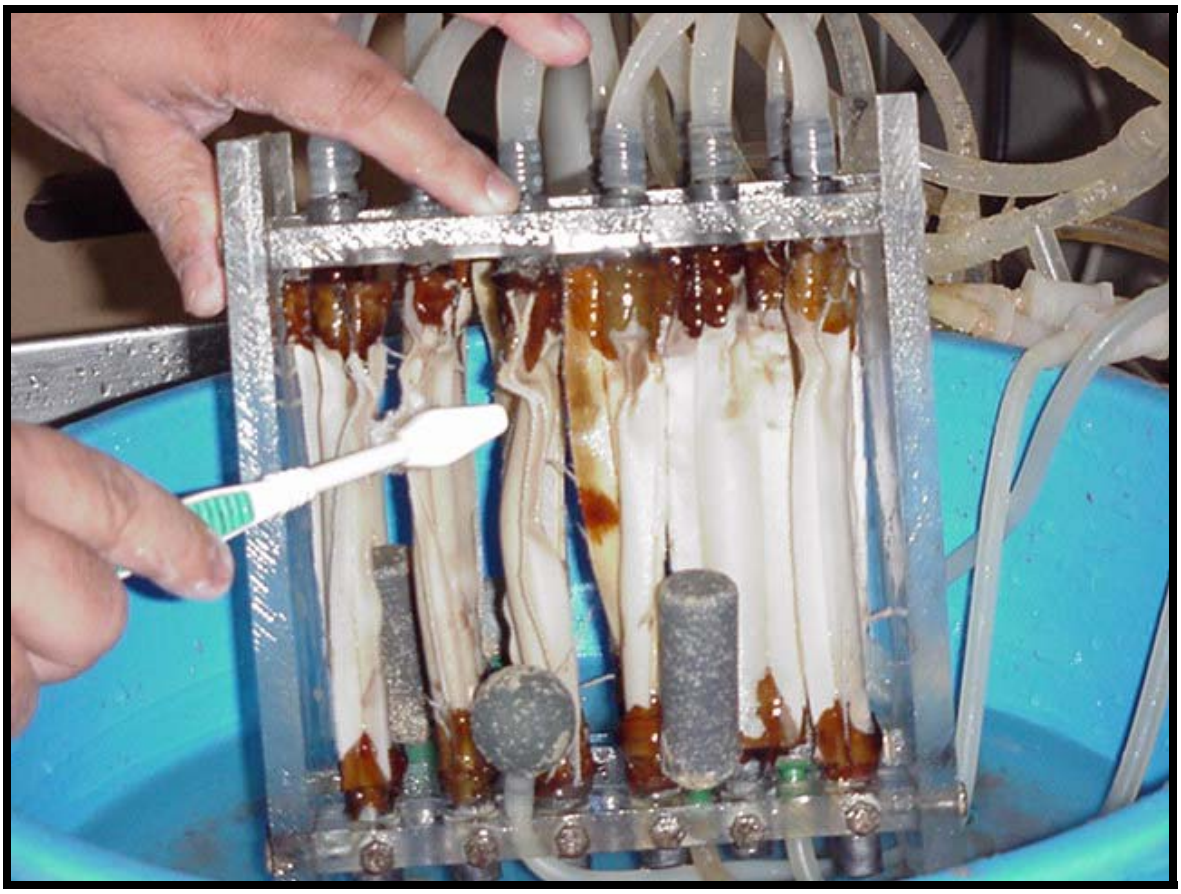

Plate 5.2: Brushing Outside Surface of Woven Fabric for Cleaning

This procedure was found to be the most effective cleaning mechanism to reinstate the membrane to its original condition. With this cleaning technique the sticky gel layer that was formed by the adsorption of the soluble matter was dislodged completely. Brushing technique also removed the sludge layer on the outer surface of the membranes as well as from the seams of the membrane. And continuous back flushing with water during the brushing helped in flushing the dislodged gel layer and sludge layer. Figure 5.7 shows the variation of the flux before and after the cleaning of the membranes using brush (Table D4). It is evident from this Figure that the after cleaning with brush, the flux improved from less than 25 l $m^{-2}hr^{-1}$ to 740 $lm^{-2}hr^{-1}$.

Along with flux restoration, the high turbidity (185 NTU) of the permeate also resulted with adoption of membrane cleaning with brushing technique. Figure 5.8 shows the variation of turbidity before and after the cleaning of the membranes using brush. However, the permeate turbidity stabilized to acceptable limits (<0.5 NTU) within one hour of operation. To maintain the effluent quality, the initial two hours permeate after every washing of the membrane with brush, was recirculated to the aeration tank.

Initially brushing outside the surface of woven fabric for cleaning was needed after every seven days. However, as the organic loading rate increased and also as mixed liquor concentration increased, the frequency of permeate flux dropping below the critical value increased, thus necessitating frequent brushing of woven fabric to restore the flux to the desired levels. During the entire experimental work, this frequency of cleaning the membrane with brush ranged from every three days to every seven days.

5.4 PERFORMANCE OF THE SUBMERGED MEMBRANE ACTIVATED SLUDGE PROCESS

As discussed above, determination of suitable pumping schedule and the appropriate membrane cleaning techniques were carried out during the one-month acclimatization period. By October 15, 2001 the system was ready for the biokinetic studies. For the next eight months the reactor was run at various organic loading rates and at different MLSS concentrations.

At the beginning of the study, the mixed liquor suspended solids concentration was kept around 3,000 mg/l, which is the same MLSS concentration usually found in conventional activated sludge process. Then, the MLSS was increased to 5,000 mg/l, 10,000 mg/l and 15,000 mg/l for determination of biokinetic coefficients. Table A1 shows the raw data for membrane performance and Biokinetic study. Figure 5.9 shows the variation of the influent substrate concentration, in terms of COD, over the biokinetic study period. As the source of substrate was glucose based synthetic wastewater, there was a greater control over influent substrate concentration, which can be clearly noticed by steady horizontal lines in Figure 5.9. The influent synthetic substrate concentration varied from 498 to 4468 mg/l as COD.

The permeate COD concentration ranged from 18 mg/l to 282 mg/l as shown in Figure 5.10. The sharp peaks in the permeate COD are due to sudden increase in influent COD concentration. The system was not monitored during the period between day 63 and day 65 because the researcher was out of station. However, the influent was supplied continuously during this period also. Figure 5.11 shows the COD removal efficiency of the SM-AS process. The COD removal efficiency ranged from 80% to over 98%.

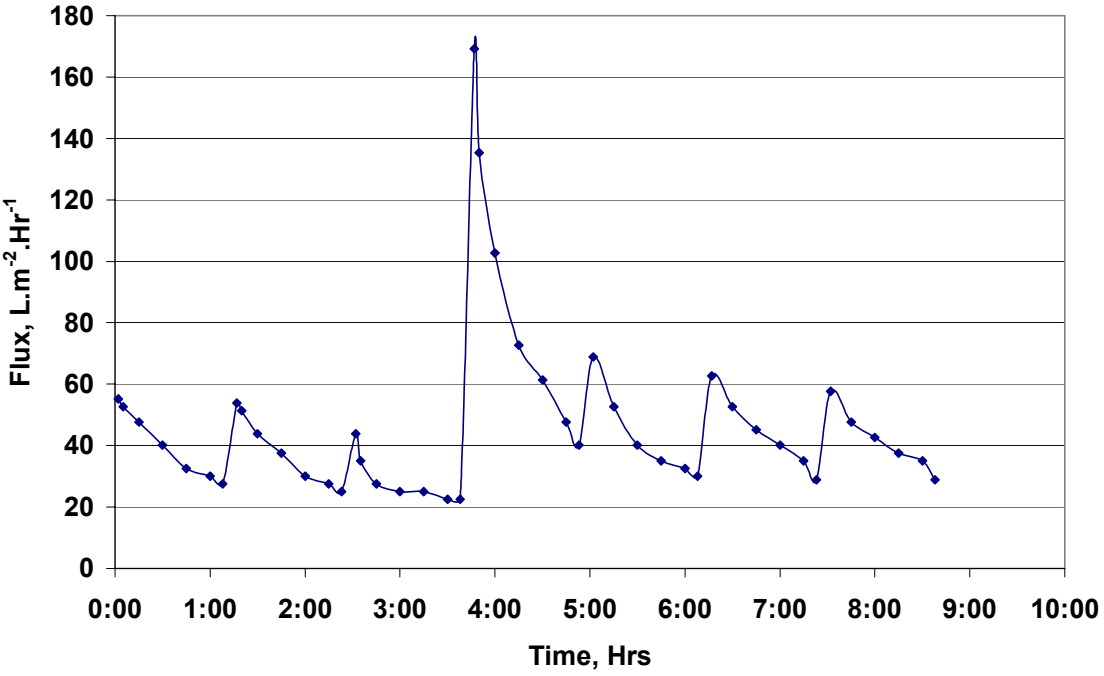

Figure 5.7: Variation of Flux Before and After Cleaning of Membranes Using Brush

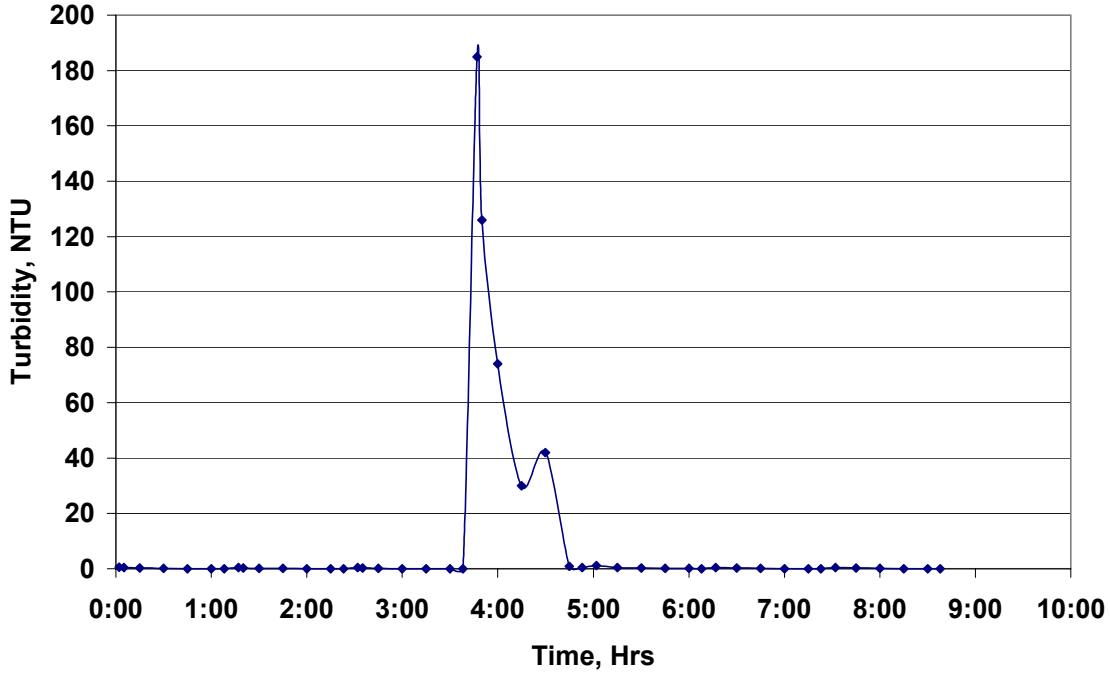

Figure 5.8: Variation of Turbidity Before and After Cleaning of Membranes Using Brush

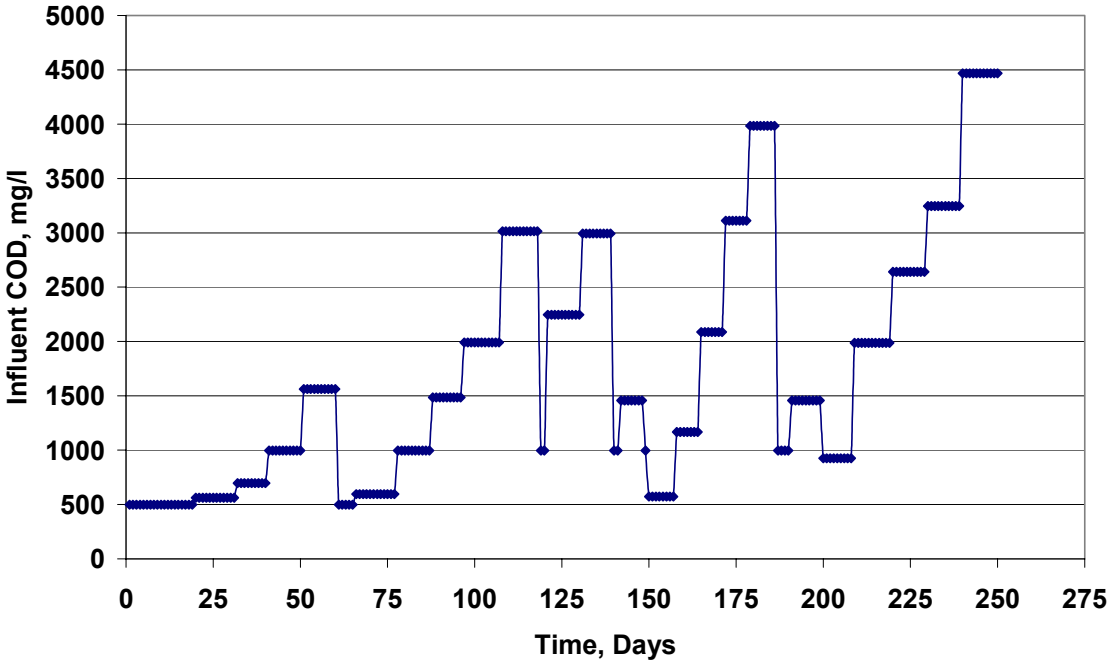

Figure 5.9: Variation of Influent COD with Time

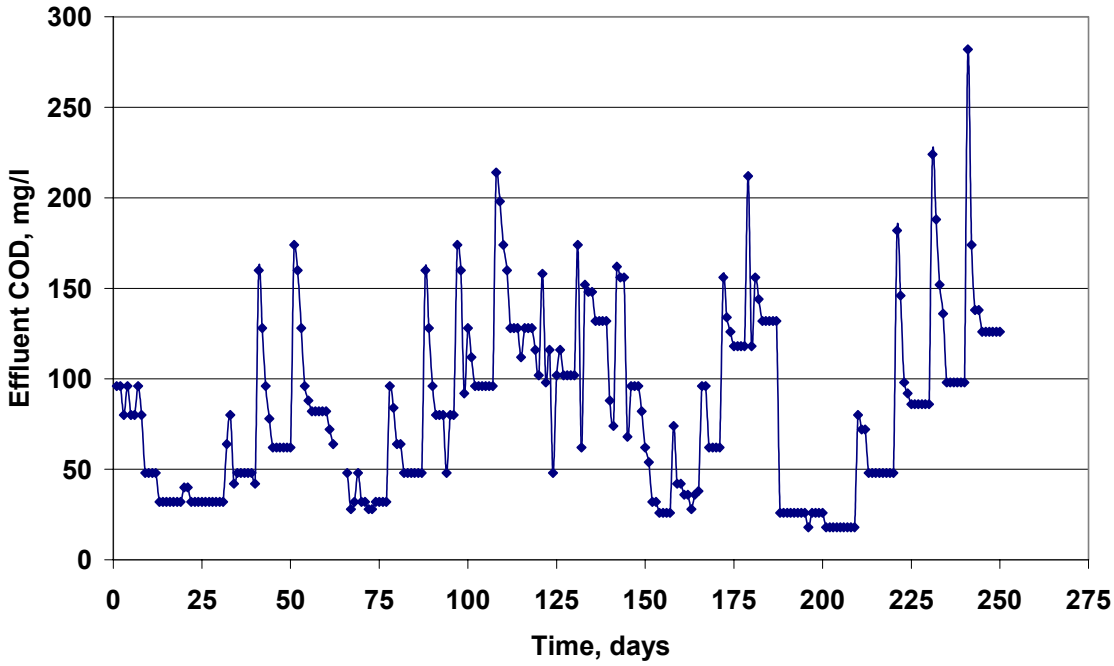

Figure 5.10: Variation of Effluent COD with Time

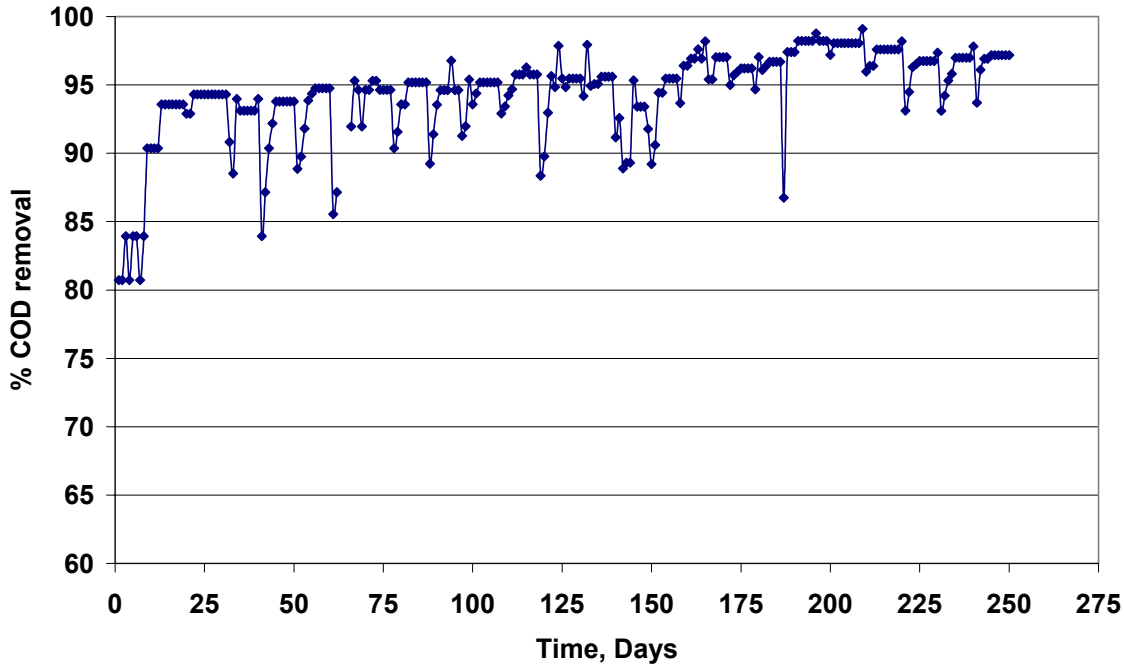

Figure 5.11: COD Removal Efficiency in the Immersed Membrane ASP

The sharp drops in COD removal efficiencies in Figure 5.11 are due to sudden increase in influent substrate COD concentration. It can be seen from the Figure that the COD removal efficiency increased with time, this is because the reactor MLSS was increasing from 3,000 mg/l to 15,000 during this period. This trend of increasing in COD removal efficiency with increase in MLSS concentration can be clearly seen in another Figure.

Permeate was collected in the permeate tank and was measured every twelve hours, giving the cumulative permeate collected. From this cumulative flux was calculated. Figure 5.12 shows the variation of cumulative flux with time. The pump "off" time was not considered for calculations of permeate flux. Thus these flux values represent cumulative flux for the system and not the true flux of the membrane.

As seen in Figure 5.12, the cumulative flux varied from 35.7 $l.m^{-2}.d^{-1}$ to 26.7 $l.m^{-2}.d^{-1}$. The frequent sharp declines seen in the Figure 5.12, are due to drop in fluxes because of pump tube rupturing. Permeate could not be collected in the permeate tank because of the tube rupturing, there by giving an erroneous cumulative volume. The tube rupturing was a regular problem faced during the experimental runs. During initial one hundred twenty days, the tube rupturing was regular but less frequent. But after this period, as the tubes became old and worn out, tube rupturing was more frequent. On 197^{th} day, all the pump's tubing was changed. After changing the tubing, the tube-rupturing problem became less frequent.

Figure 5.13 shows variation of hydraulic retention time (θ) during the study period. Though the system was designed for the average hydraulic retention time of twelve hours, at higher mixed liquor suspended solids concentrations twelve hours HRT could

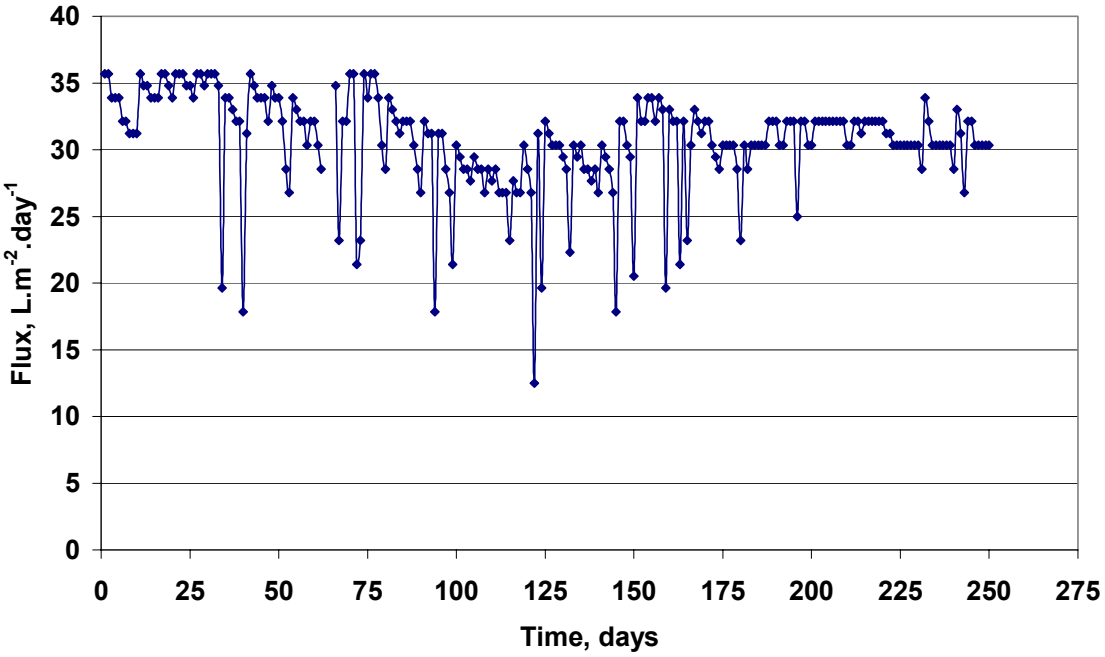

Figure 5.12: Variation of Cumulative Flux with Time

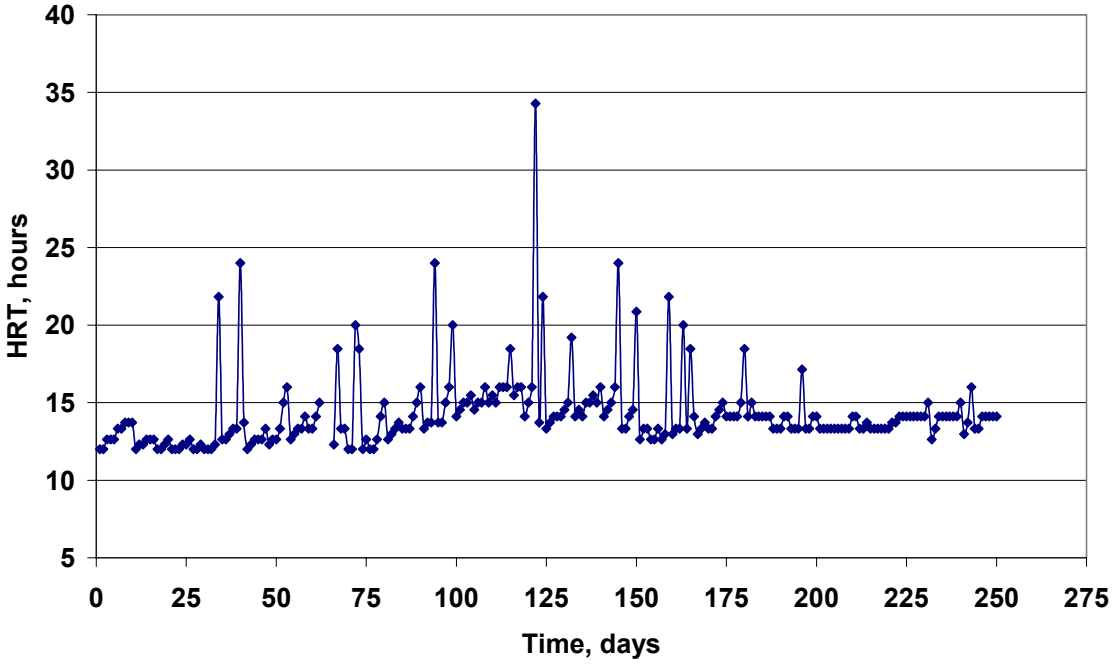

Figure 5.13: Variation of Hydraulic Retention Time

not be achieved. As the influent substrate concentration increased the mean permeate flux dropped to lower than 35 $l.m^{-2}.d^{-1}$. This flux could not be enhanced because of pump limitations.

For biokinetic coefficient determination, the system was operated at hydraulic retention time (HRT) varying from twelve hours to fifteen hours. The sharp peaks at regular intervals in the Figure 5.13 were due to increase in HRT because of pump tubing. The ruptured tube was replaced as and when the rupture was detected, and the HRT came back to within designed limits.

Though for determination of biokinetic coefficients, HRT is an important parameter that can be varied, but because of the limitations imposed by the microfiltration unit providing a higher filtration rate, HRT could not be used as a controlling parameter. Hence the HRT was within a narrow range of 12 to 15 hours duration.

The mixed liquor suspended solids concentration of the reactor was measured twice a day. One measurement was to monitor the MLSS concentration present in the reactor and then calculate the volume of MLSS to be wasted in order to keep the suspended solids at the desired concentration levels. The other measurement was taken after the wasted volume of the mixed liquor was replaced with tap water in order to check the remaining MLSS concentration in the reactor. Figure 5.14 shows the variation of MLSS concentration in the reactor before and after wasting during the period of kinetic studies. In the above mentioned Figure the peaks show the MLSS concentration before wasting of the biomass and dips show the MLSS concentration after wasting of the biomass.

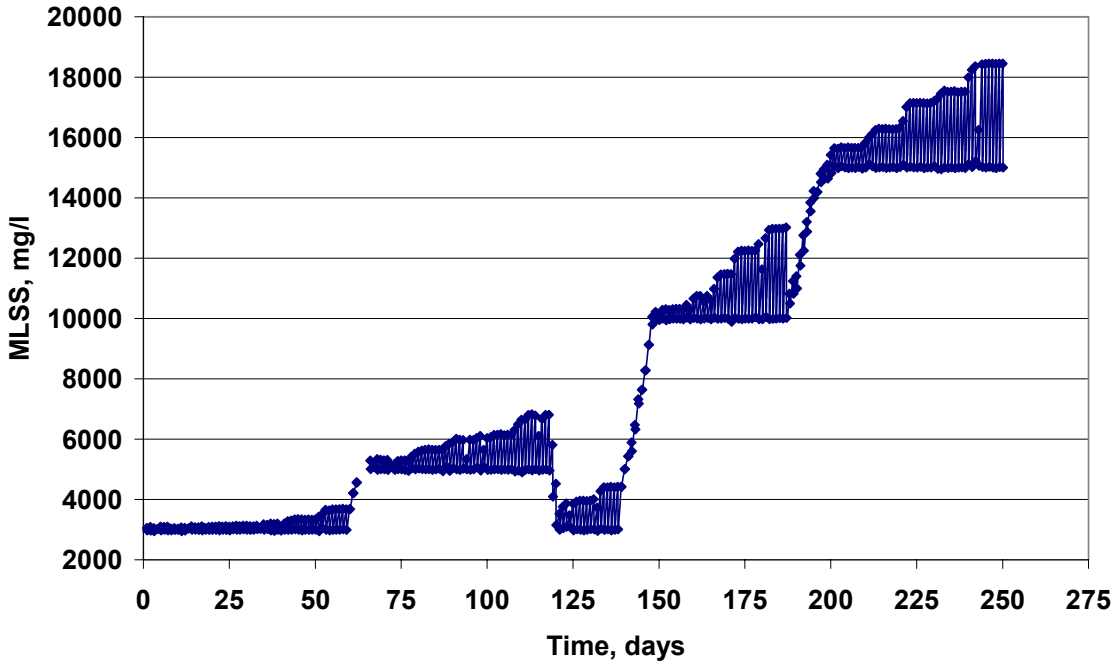

Figure 5.14: Variation of MLSS Concentration with Time

Before wasting of the mixed liquor, the MLSS concentration varied from 3060 mg/l to 18,450 mg/l and after wasting of the biomass, the MLSS concentration varied from 2,950 mg/l to 18,450 mg/l., during the period of study as shown in Figure 5.14. The crisscross shape of the curve is because of the daily wasting of the bio solids

In order to estimate the volatile portion of the biomass, mixed liquor volatile suspended solid concentration was measured daily before sludge wasting. Figure 5.15 shows the variation of MLVSS/MLSS ratio with time during the period of biokinetic study. The MLVSS/MLSS ratio varied from 68.53% to 90.5%. Initially, the MLVSS/MLSS ratio was very high, around 87% as expected, as synthetic substrate was used as influent, which was free from fixed solids. With time, volatile content of the mixed liquor declined because of accumulation of non volatile compounds in the reactor.

During the period of kinetic study, the organic loading rate which is also called F/M ratio (Food to Microorganism ratio) varied from 0.11 to 1.19 kg COD/kg MLSS per day as shown in Figure 5.16. The organic loading rate was varied to give different sludge loading rates for a given MLSS concentration. With these organic loading rates, foaming was never observed during the biokinetic study phase. The peaks and sudden drop in organic loading rates are due to sharp changes in the influent substrate concentration.

The dissolved oxygen concentration, which was measured frequently using a DO probe, was never below 6.2 mg/l. This shows that the aeration provided in the reactor was in excess of dissolved oxygen requirement. However, this air supply was needed to keep the reactor contents in suspension. A maximum dissolved oxygen content of 7.1 mg/l was recorded during the experimental run. Dissolved oxygen was not a parameter under

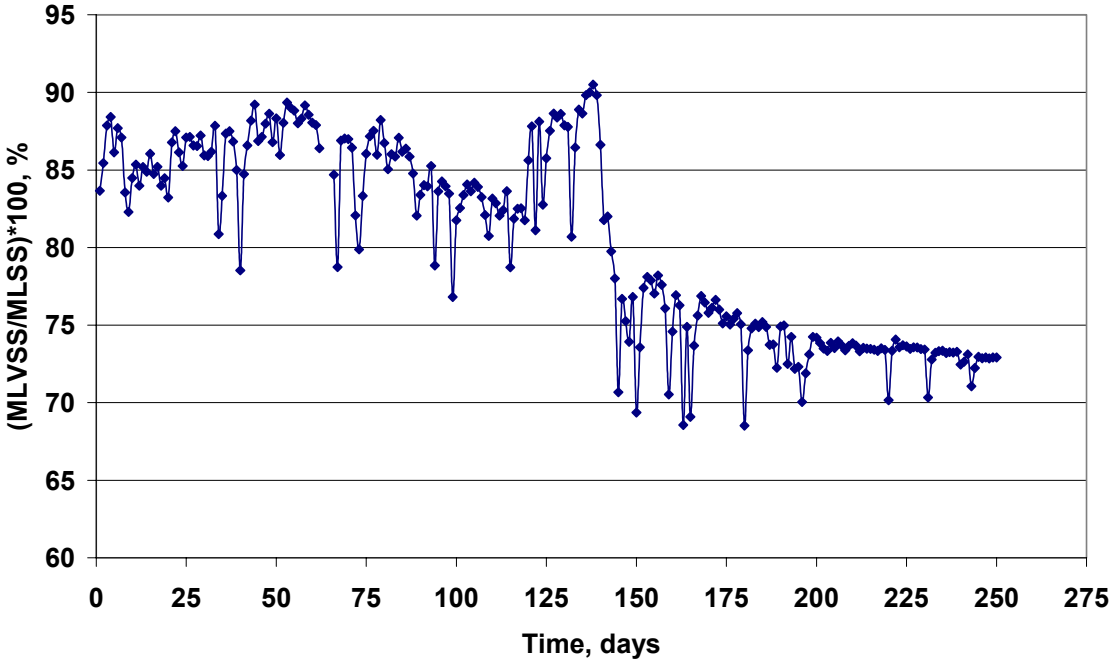

Figure 5.15: Variation of MLVSS/MLSS with Time

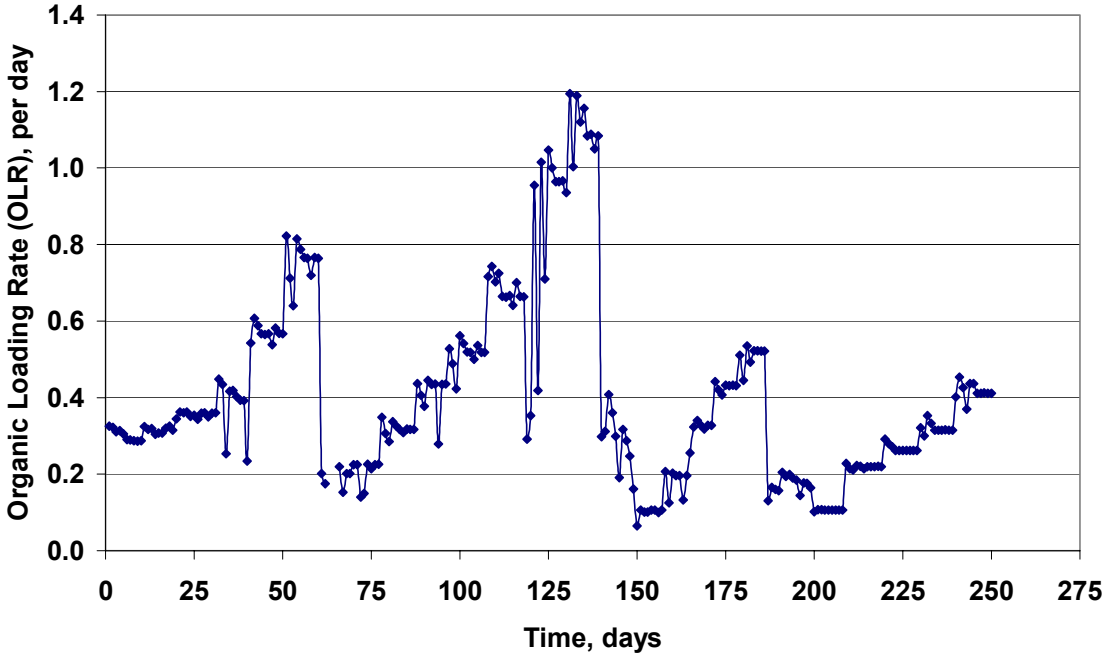

Figure 5.16: Variation of Organic Loading Rate (OLR)

consideration for the biokinetic study

Sludge retention time (SRT) was used as one of the parameter to control the growth rate of biomass for the determination of kinetic coefficients. Figure 5.17 shows the variation of sludge retention time during the period of kinetic coefficients study. The sludge retention time varied from two days to seventy-four days during the study period. The sharp peaks and falls in the graph were due to the intermittent wasting of the mixed liquor to maintain the reactor MLSS concentration at desired levels. The discontinuity of the trend line is due to the fact that sludge was not wasted on that day. It can be easily seen in Figure 5.17 that, initially, SRT was high. As the MLSS concentration increased, SRT decreased. Also as the influent substrate concentration increased, SRT decreased.

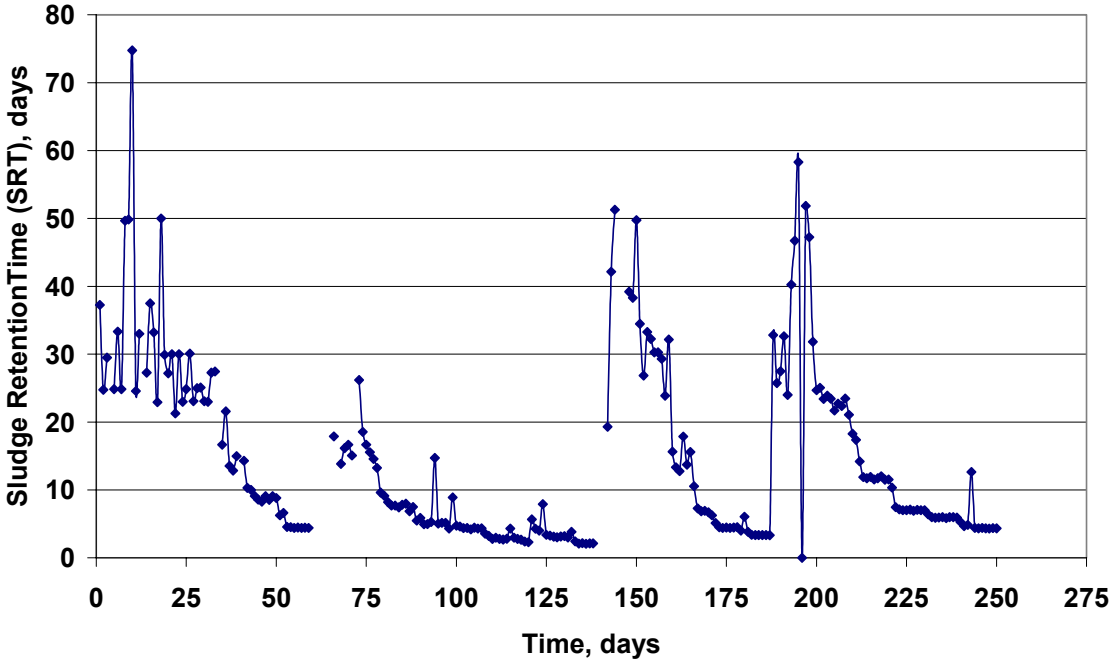

Figure 5.17: Variation of Sludge Retention Time (SRT)

5.5 DETERMINATION OF BIOKINETIC COEFFICIENTS

5.5.1 Development of Kinetic Model equations

The basic equations that describe the interaction between the growth of microorganisms and utilization of the growth-limiting substrate in the activated sludge process are based on the Monod (1949) equations. The Monod model is still the most commonly and widely used model for the study of biokinetic coefficients. This model was accepted by the IAWPRC task group (Henze et al., 1986) as the fundamental basis for the development of the activated sludge model.

The microorganisms require substrate for three main functions:

i. To synthesize the new cell material

ii. To synthesize the extra cellular products and

iii. To provide the energy necessary to drive the synthetic reaction and maintain concentrations of materials within the cell which are different from those in the environment.

In both batch and continuous culture systems the rate of growth of bacterial cells can be defined by the following relationship:

$$r_g = \mu X \tag{5.1}$$

Where r_g = rate of bacterial growth, mass/unit volume. time

μ = specific growth rate, $time^{-1}$

X = concentration of microorganisms, mass/unit volume

Because dX/dt = r_g for the batch culture, the following relationship is also valid for a batch reactor:

$$\frac{dX}{dt} = \mu X \tag{5.2}$$

The effect of a limiting substrate or nutrient can often be defined adequately using the following expression proposed by Monod (1949);

$$\mu = \mu_m \frac{S}{K_s + S} \tag{5.3}$$

Where, t = time

μ_m = maximum specific growth rate, $time^{-1}$

S = concentration of growth limiting substrate surrounding the biomass, mass/unit volume

K_s = saturation constant which is numerically equal to the substrate concentration at $\mu = 0.5\ \mu_m$, mass/ unit volume

Substituting the value of μ from Equation 5.3 in Equation 5.1, the resulting expression for the rate of growth is:

$$r_g = \frac{\mu_m X S}{K_s + S} \tag{5.4}$$

In the batch and continuous growth culture a system, a portion of the substrate is converted to new cells and portion is oxidized to inorganic and organic end products. The relationship between the mass of bacteria produced and the mass of organic substrate removed is quantified by a coefficient known as yield coefficient, Y, and numerically expressed as:

$$Y = \frac{dX / dt}{dS / dt} \tag{5.5}$$

The yield coefficient is usually assumed to be constant for a given biological process treating a specific waste. Also yield depends on

- Various physical parameters of cultivation,
- The degree of polymerization of the substrate,
- Pathways of metabolism,
- The growth rate, and
- The oxidation state of the carbon source and nutrient elements.

The following relationship has been developed between the rate of substrate utilization and the rate of growth:

$$r_g = -Y\, r_{su} \tag{5.6}$$

Where r_{su} = substrate utilization rate, mass/unit volume. Time

In bacterial systems used for wastewater treatment, the distribution of cell ages is such that not all the cells in the system are in the log-growth phase. Consequently, the expression for the rate of growth must be corrected to account for the energy require for cell maintenance. Other factors, such as death and predation, must also be considered. Usually, these factors are lumped together, and it is assumed that the decrease in cell mass caused by them is proportional to the concentration of organism present. This decrease is known as endogenous decay, r_d, and it can be formulated as:

$$r_d = -k_d X \tag{5.7}$$

Where k_d = endogenous decay coefficient, $time^{-1}$

The growth of the biomass in the process can be expressed as:

$$\frac{dX}{dT} = \mu\ X - k_d X \tag{5.8}$$

Combining equations 5.1 and 5.5 gives:

$$\frac{dS}{dt} = \mu \frac{X}{Y} \tag{5.9}$$

Rearranging equation 5.9 and substituting in equation 5.8 we get:

$$\frac{dX}{dt} = Y \frac{dS}{dt} - k_d X \tag{5.10}$$

Rearranging equation 5.10 gives:

$$\mu = U\ Y - k_d \tag{5.11}$$

Where, U = specific substrate utilization rate, time^{-1}, and is represented by:

$$U = \frac{Q(S_o - S)}{V\ X} \tag{5.12}$$

The above equations (from 5.1 to 5.12) when combined, form the basis of the mathematical model for the submerged membrane activated sludge process.

Figure 5.18 shows the schematic diagram of the SM-AS system. The model is developed with the following assumptions:

i. the reactor is completely mixed (mixing was provided by the stone aerators provided at the bottom of the filtration unit as well as at the near end of the tank),
ii. influent substrate concentration remains constant (this was achieved by choosing synthetic influent as substrate),
iii. no microbial solids are contained in the influent substrate,

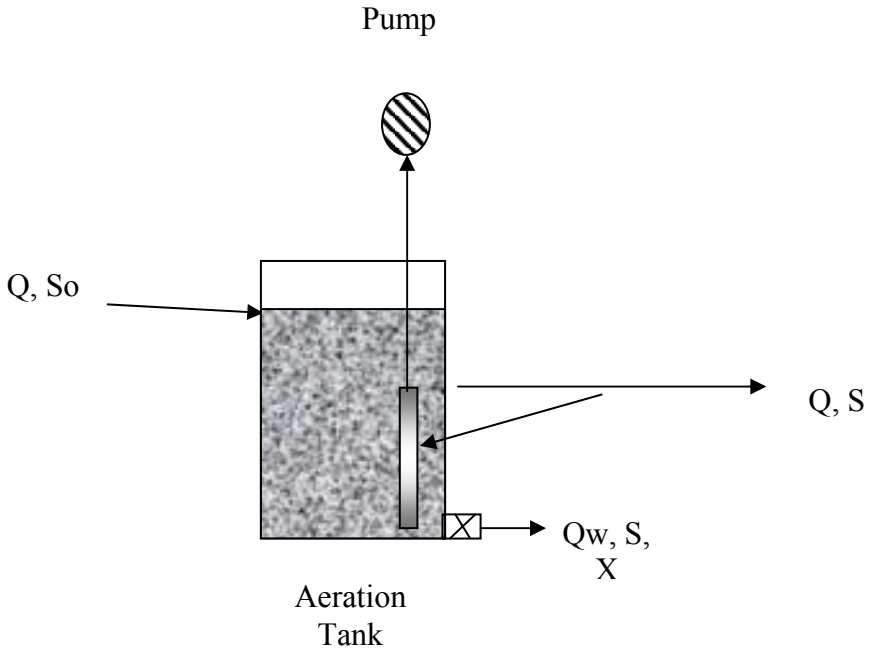

Figure 5.18 Completely Mixed SM-AS Reactor

iv. the volume of the reactor is constant (the inflow rate is equal to the permeate flux), this was achieved by means of a mechanical float,

v. complete rejection of the MLSS (no biomass is allowed to come out with the permeate),

vi. substrate is not rejected (since the membrane as a high molecular weight-cutoff and the glucose has a low molecular weight), and steady state conditions prevail throughout the system.

The rate equations describing the performance of the system are the mass balance equations of both the biomass and substrate. These can be expressed as follows:

Biomass balance:

$$\begin{bmatrix}\text{Rate of change of}\\ \text{biomass in the reactor}\end{bmatrix} = \begin{bmatrix}\text{Rate of increase}\\ \text{due to growth}\end{bmatrix} - \begin{bmatrix}\text{Rate of loss due to}\\ \text{endogenous respiration}\end{bmatrix} - \begin{bmatrix}\text{Delibarate}\\ \text{wastage}\end{bmatrix}$$

The symbolic representation of the above statement is:

$$V\frac{dX}{dt} = \mu\ X\ V - k_d\ X\ V - Q_w\ X \qquad (5.13)$$

Where, V = reactor volume, L

X = biomass concentration in the reactor, mg/l

μ = specific growth rate, day^{-1}

k_d = biomass decay coefficient, day^{-1}

Q_w = wastage flow rate, l/sec

For steady state conditions, dX/dt = 0;

Hence, the equation 5.13 changes to,

$$\mu = k_d + \frac{Q_w}{V} \qquad (5.14)$$

Since the solid retention time (SRT) is defined as:

$$SRT = \frac{Total\ mass\ of\ organisms\ in\ the\ reactor}{Total\ mass\ of\ organisms\ leaving\ the\ sysrem\ per\ day}$$

Then,

$$SRT = \frac{V\ X}{Q_w\ X} = \frac{V}{Q_w} \tag{5.15}$$

Substituting equation 5.15 in equation 5.14, we get:

$$\mu = k_d + \frac{1}{SRT} \tag{5.16}$$

Substituting equation 5.3 in equation 5.16 yields the steady state for substrate concentration in the reactor:

$$S = \frac{K_s\left(\frac{1}{SRT} + k_d\right)}{\mu_m - \left(\frac{1}{SRT} + k_d\right)} \tag{5.17}$$

Substrate balance:

$$\begin{bmatrix}\text{The rate of}\\ \text{change of}\\ \text{substrate in}\\ \text{the reactor}\end{bmatrix} = \begin{bmatrix}\text{Rate of input}\\ \text{of the feed}\\ \text{substrate}\end{bmatrix} - \begin{bmatrix}\text{Rate of}\\ \text{removal due}\\ \text{to biomass}\\ \text{utilization}\end{bmatrix} - \begin{bmatrix}\text{Rate of}\\ \text{removal}\\ \text{due to}\\ \text{washout}\end{bmatrix} - \begin{bmatrix}\text{Substrate lost}\\ \text{during delibarate}\\ \text{wastage}\\ \text{of biomass}\end{bmatrix}$$

The mathematical description of the above statement can be written as:

$$V\frac{dS}{dt} = Q\ S_o - \mu\frac{X\ V}{Y} - S(Q - Q_w) - Q_w S \tag{5.18}$$

At steady state, dS/dt = 0;

Therefore, above equation changes to:

$$\frac{Q}{V}(S_o - S) = \mu\frac{X}{Y} \tag{5.19}$$

Substituting equation 5.16 into equation 5.19 gives the biomass concentration at steady state condition:

$$X = Y\frac{Q}{V}\frac{(S_o - S)}{k_d + \frac{1}{SRT}} \tag{5.20}$$

5.5.2 Determination of Kinetic Coefficients

The purpose of studying the kinetic coefficients was to obtain information on the rate of cell growth and consumption of substrate. This enabled the required volume of the reactor to be calculated and simulation of the system can be used for process control. The kinetic coefficients of a biological system have generally been determined experimentally using either continuous flow, completely mixed or batch lab-scale reactors.

In continuous flow completely mixed reactor, the determination of the kinetic coefficients is usually achieved by collecting data from lab-scale or pilot-plant experiments. Operating the system at various hydraulic retention times (HRT) and/or at various sludge retention times, and by allowing (at each adapted stage or HRT or SRT), a steady state condition to prevail. Accurate measurements of the biomass and permeate substrate concentration are then recorded. The parameters such as K_s, μ, Y and k_d can be determined through linearization of equations 5.17 and 5.20 as follows:

To determine the kinetic coefficients, k_d and Y, rearranging equations 5.20 gives:

$$\frac{Q}{VX}(S_o - S) = \frac{1}{Y}\frac{1}{SRT} + \frac{k_d}{Y} \tag{5.21}$$

To determine the kinetic coefficients, μ_m and K_s, equation 5.17 can be re arranged as:

$$\frac{SRT}{1+(SRT\ k_d)} = \frac{K_s}{\mu_m}\left(\frac{1}{S}\right) + \frac{1}{\mu_m} \qquad (5.22)$$

If equation 5.21 is plotted as $Q(S_o\text{-}S)/VX$ versus 1/SRT, then from the slope and the intercept, it is possible to determine, the kinetic coefficients, k_d and Y. Substituting the obtained value of k_d in equation 5.22 and plotting $SRT/[1+(SRT\ k_d)]$ versus 1/S, then from the slope and the intercept it is possible to determine, the kinetic coefficients, μ_m and K_s.

The studies of the kinetic coefficients in the SM-AS process were carried out in a similar fashion to that outlined above. However, due to the limitation imposed by the microfiltraion unit providing a higher filtration flux, the hydraulic retention time could not be used as a parameter to control the growth rate of the biomass, and sludge retention time was adopted instead. This was achieved by operating the unit at various organic loading rates and also by wasting the various volumes of biomass from the system.

The concentration of the biomass was kept constant, by wasting mixed liquor once a day. A steady state condition was achieved when fairly constant biomass growth and filtrate COD were attained. Initially, the biomass concentration in the aeration tank is kept at 3000 mg/l. Since the biomass was already acclimatized to the synthetic glucose based feed, the first steady state condition was achieved after only twenty two days from the start of the unit operation. The steady state of maintained for about nine days, measuring carefully increase in biomass and permeate quality.

It was then decided to increase the organic loading rate (OLR) from 0.37 kg COD/ kg MLSS day to 0.60 kg COD/ kg MLSS day, and second steady state condition was

achieved after two weeks from the changing of the OLR. The steady state condition was maintained for about six days, after which it was decided to increase the OLR to 0.84 kg COD/ kg MLSS day to achieve a third steady state condition after six days from changing the OLR. The unit was operated at this stage for five days.

Initially, it was thought to have three steady state conditions only for a given MLSS concentration; hence the MLSS concentration was increased from 3000 mg/ to 5000 mg/l. This increase was achieved in five days without sludge wastage. However, it was felt necessary to have additional plotting points to calculate kinetic coefficients more accurately. Hence once again the MLSS concentration was changed from 5000 mg/l to 3000 mg/l on day 120 and fourth steady state condition was established at an OLR of 1.07 kg COD/ kg MLSS day. The steady state condition was maintained for four days, and then OLR was increased to 1.29 kg COD/ kg MLSS day.

Table 5.1 shows the steady state data for MLSS concentration of 3000 mg/l. A linear regression on the points in Table 5.1 in accordance with equations 5.21 and 5.22 was used. The plots are shown in Figures 5.19 and 5.20 respectively, and the kinetic coefficients were: Y = 0.487 mg/mg, k_d = 0.151 day^{-1}, μ_m = 1.28 day^{-1} and K_s = 289 mg COD/l. An organic loading rate of 1.07 kg COD/ kg MLSS day gave the maximum COD removal efficiency (95.6% COD removal) at 3000 mg/l MLSS concentration.

When the system was operated at an MLSS concentration of 5000 mg/l from 66^{th} day to 118^{th} day the OLR was varied from 0.23 kg COD/ kg MLSS day to 0.77 kg COD/ kg MLSS day. The results at the steady state conditions are shown in Table 5.2. Figure 5.21.

Table 5.1: Steady State Data at MLSS 3000 mg/l

Steady State	Q	HRT	Xincr.	Xavg	So	S	OLR	SRT	Q(So-S)/(VX)	SRT/(1+SRT*kd)
Period	l/day	hrs	mg/l	mg/l	mg/l	mg/l	1/day	days	1/day	day
22 - 31	40	12	3120	3060	562	32	0.37	25	0.35	5.24
45 - 50	38	12.63	3340	3170	996	62	0.6	8.82	0.56	3.78
56 - 60	36	13.33	3680	3340	1562	82	0.84	4.41	0.8	2.65
127 - 130	34	14.12	3960	3560	2246	102	1.07	3.13	1.02	2.12
136 - 139	32	15	4420	3710	2994	132	1.29	2.11	1.23	1.6

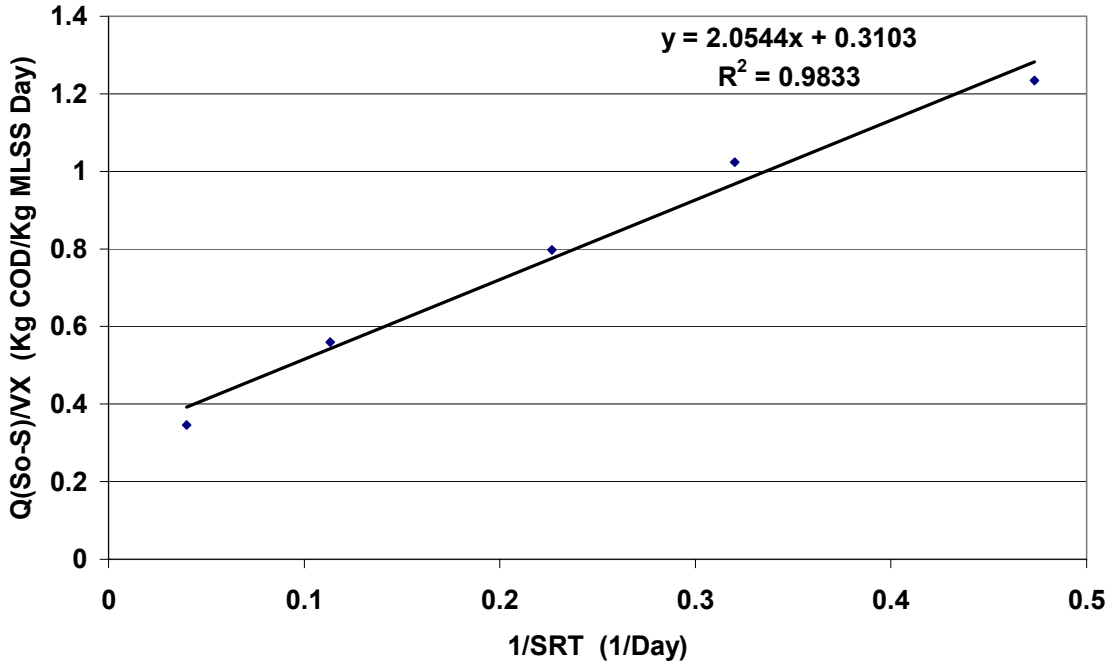

Figure 5.19: Determination of Y and k_d at MLSS 3000 mg/l

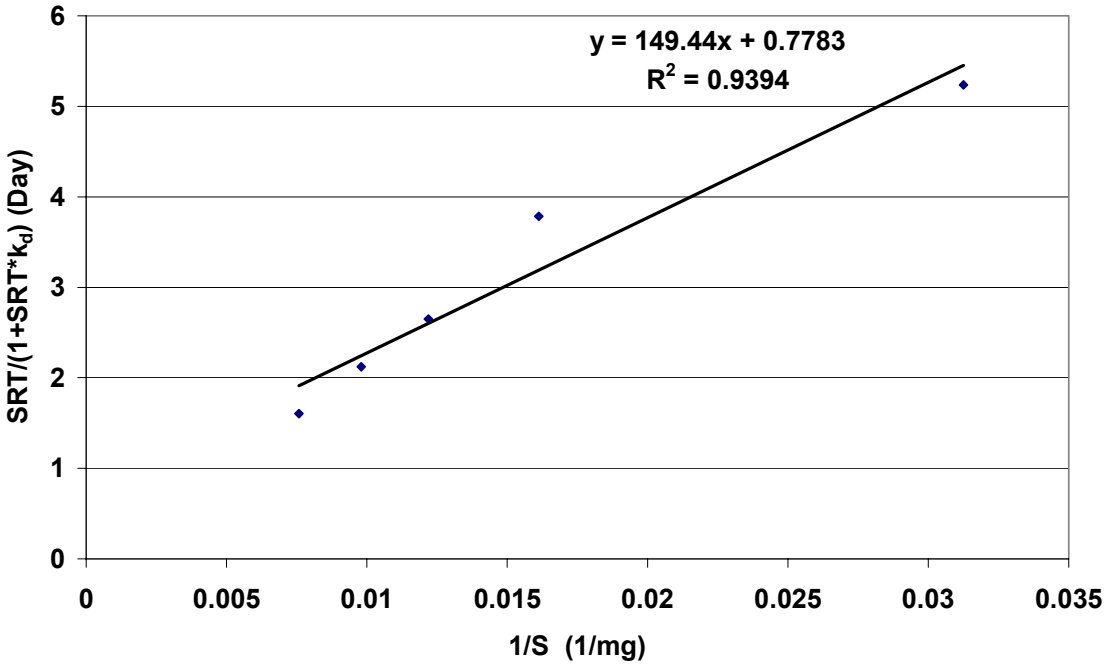

Figure 5.20: Determination of μ_m and K_s at MLSS 3000 mg/l

Table 5.2: Steady state data at MLSS 5000 mg/l

Steady State	Q	HRT	Xincr.	Xavg	So	S	OLR	SRT	Q(So-S)/(VX)	SRT/(1+SRT*kd)
Period	l/day	hrs	mg/l	mg/l	mg/l	mg/l	1/day	days	1/day	day
70 - 76	40	12	5300	5150	596	32	0.23	16.67	0.22	8.2
82 - 87	36	13.33	5650	5325	996	48	0.34	7.69	0.32	5.21
95 - 96	35	13.71	5980	5490	1486	80	0.47	5.1	0.45	3.88
102 - 107	32	15	6150	5575	1992	96	0.57	4.35	0.54	3.42
112 - 118	30	16	6800	5900	3014	128	0.77	2.78	0.73	2.37

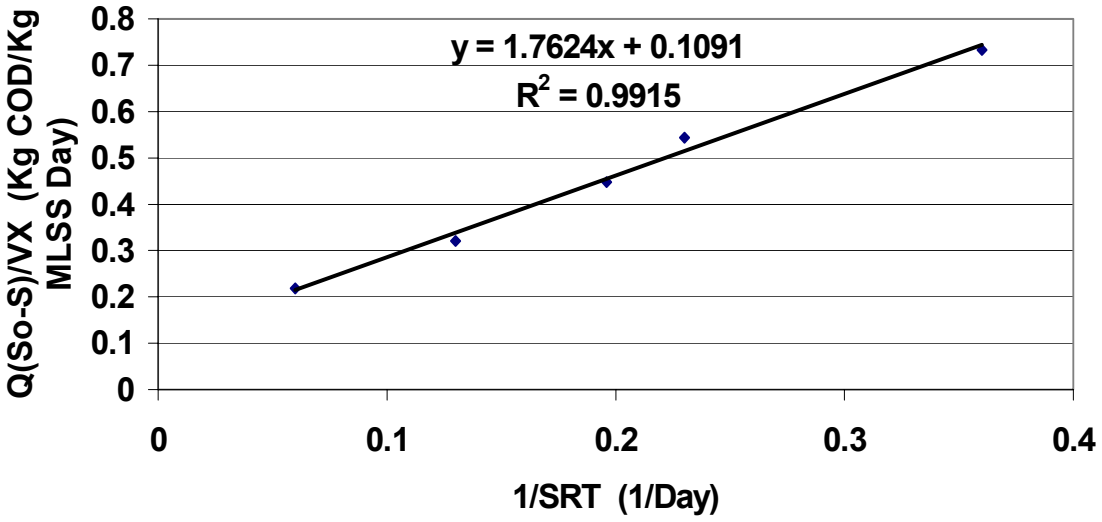

Figure 5.21: Determination of Y and k_d at MLSS 5000 mg/l

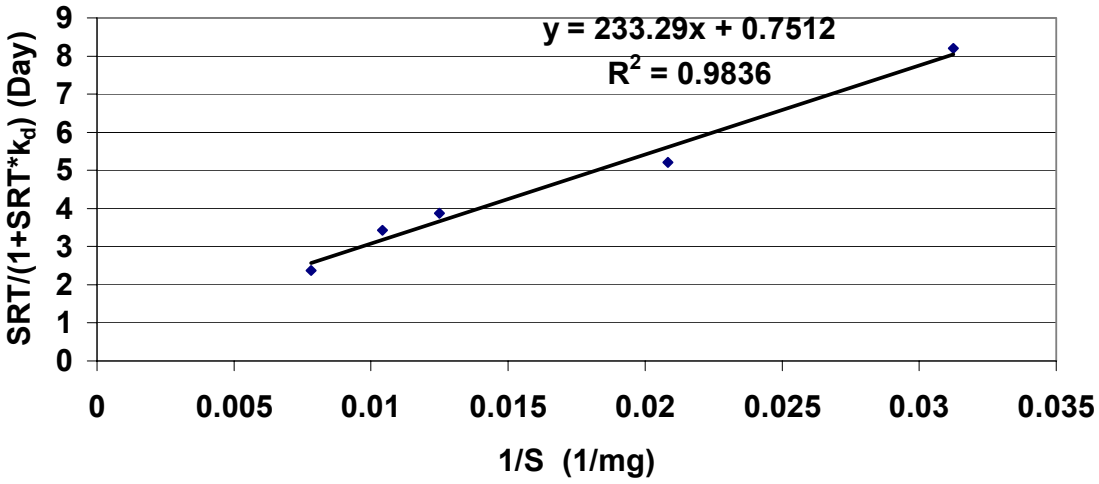

Figure 5.22: Determination of μ_m and K_s at MLSS 5000 mg/l

and 5.22 show the plots of the steady state data obtained from table 5.2 for determination of kinetic coefficients.

The values of the kinetic coefficients were: Y = 0.567 mg/mg, k_d = 0.062 day^{-1}, μ_m = 1.398 day^{-1} and K_s = 326 mg COD/l. During this runs, variation of OLR did not had much influence on the COD removal efficiency. The COD removal efficiencies were in the narrow range of 94 – 95%COD removal

After this, the concentration of biomass was increased to 10 000 mg/l by 148^{th} day. In this run, an OLR of 0.11 kg COD/ kg MLSS day was applied initially and later increased to 0.59 kg COD/ kg MLSS day. The steady state data are presented in Table 5.3 and plots of the steady state points for the determination of the growth kinetics are shown in Figures 5.23 and 5.24. The values of the kinetic coefficients for the 10 000 mg/l of mixed liquor suspended solids were: Y = 0.571 mg/mg, k_d = 0.037 day^{-1}, μ_m = 5.520 day^{-1} and K_s = 1967 mg COD/l. Maximum COD removal efficiency (97%) was achieved at an OLR of 0.32 kg COD/ kg MLSS day.

After finishing of steady state analysis at MLSS concentration of 10 000 mg/l, the mixed liquor concentration was raised to 15 000 mg/l in twelve days with daily wasting of some amount of sludge. The first steady state condition at 15 000 MLSS concentration was achieved on 204^{th} day of operation and steady state lasted for five days with an OLR of 0.11 kg COD/ kg MLSS day. The COD removal was 98 % and was the highest COD removal achieved during entire steady state runs. After this, as the OLR increased, COD removal efficiency decreased, although efficiency remained above 95% COD removal.

Table 5.3: Steady State Data at MLSS 10000 mg/l

Steady State	Q	HRT	Xincr.	Xavg	So	S	OLR	SRT	Q(So-S)/(VX)	SRT/(1+SRT*kd)
Period	l/day	hrs	mg/l	mg/l	mg/l	mg/l	1/day	days	1/day	day
154 - 157	38	12.63	10320	10160	574	26	0.11	31.25	0.1	14.71
161 - 164	36	13.33	10750	10375	1168	36	0.2	13.33	0.2	9.01
168 - 171	36	13.33	11470	10735	2086	62	0.35	6.8	0.35	5.46
175 - 178	34	14.12	12260	11130	3112	118	0.48	4.42	0.47	3.82
183 - 186	34	14.12	12980	11490	3984	132	0.59	3.36	0.57	2.99

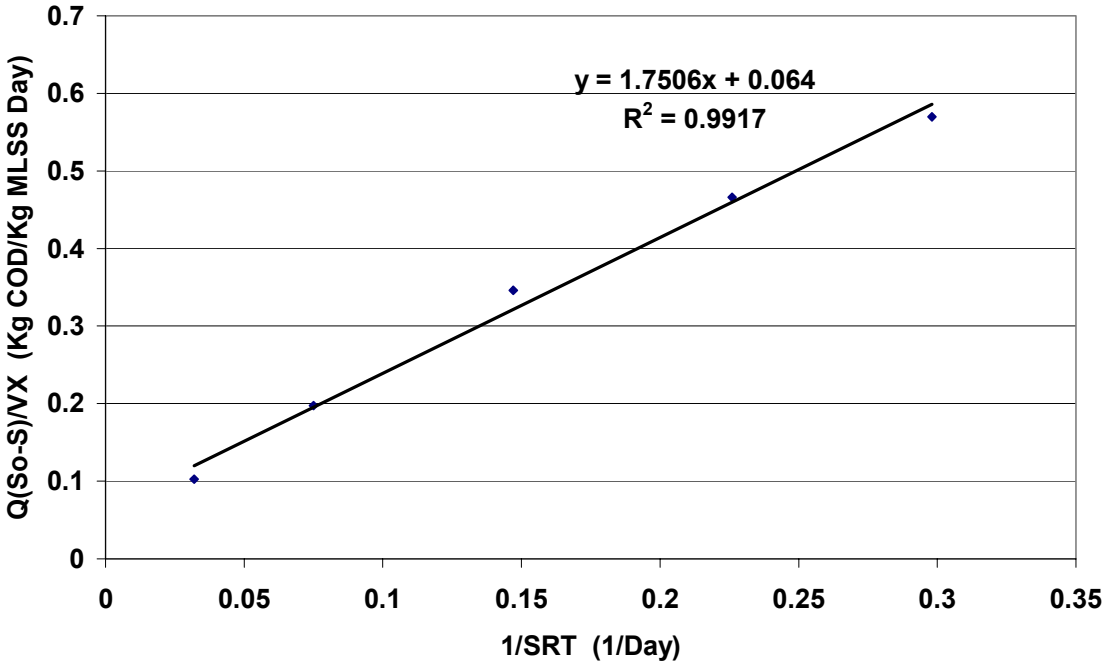

Figure 5.23: Determination of Y and k_d at MLSS 10000 mg/l

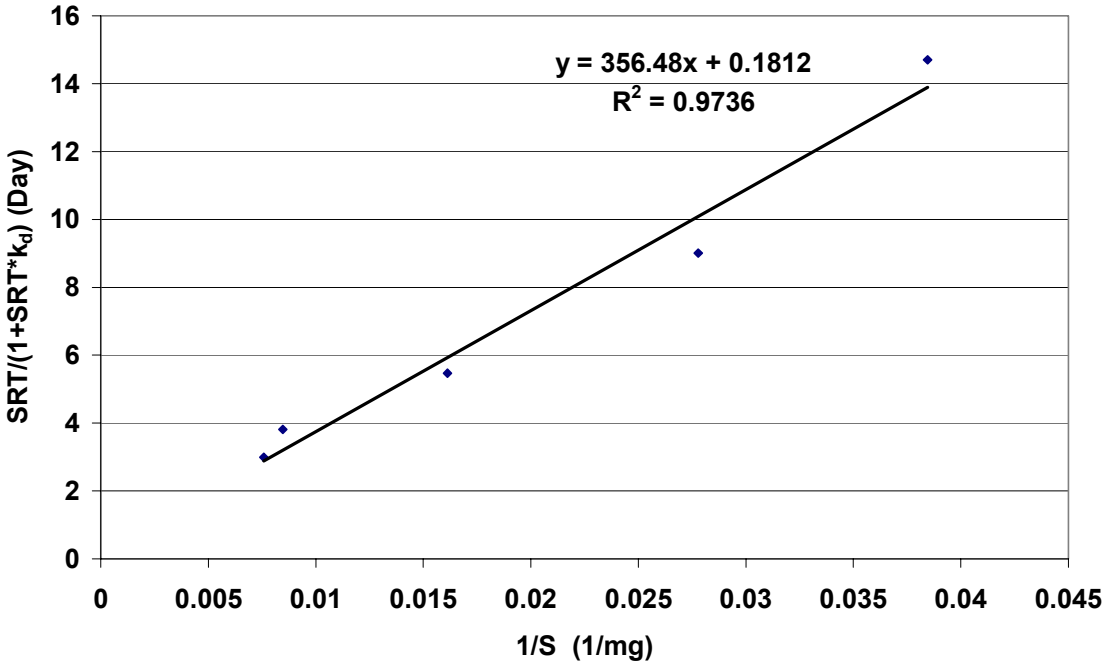

Figure 5.24: Determination of μ_m and K_s at MLSS 10000 mg/l

Table 5.4: Steady state data at MLSS 15000 mg/l

Steady State	Q	HRT	Xincrease	Xavg	So	S	OLR	SRT	Q(So-S)/(VX)	SRT/(1+SRT*kd)
Period	l/day	hrs	mg/l	mg/l	mg/l	mg/l	1/day	days	1/day	day
204-208	36	13.3	15660	15330	926	32	0.11	22.73	0.1	14.27
215-219	36	13.3	16280	15640	1988	52	0.23	11.72	0.22	8.97
225-229	34	14.1	17140	16070	2642	86	0.28	7.01	0.27	5.93
235-239	34	14.1	17520	16260	3246	92	0.34	5.95	0.33	5.15
246-250	34	14.1	18450	16725	4468	106	0.45	4.35	0.44	3.9

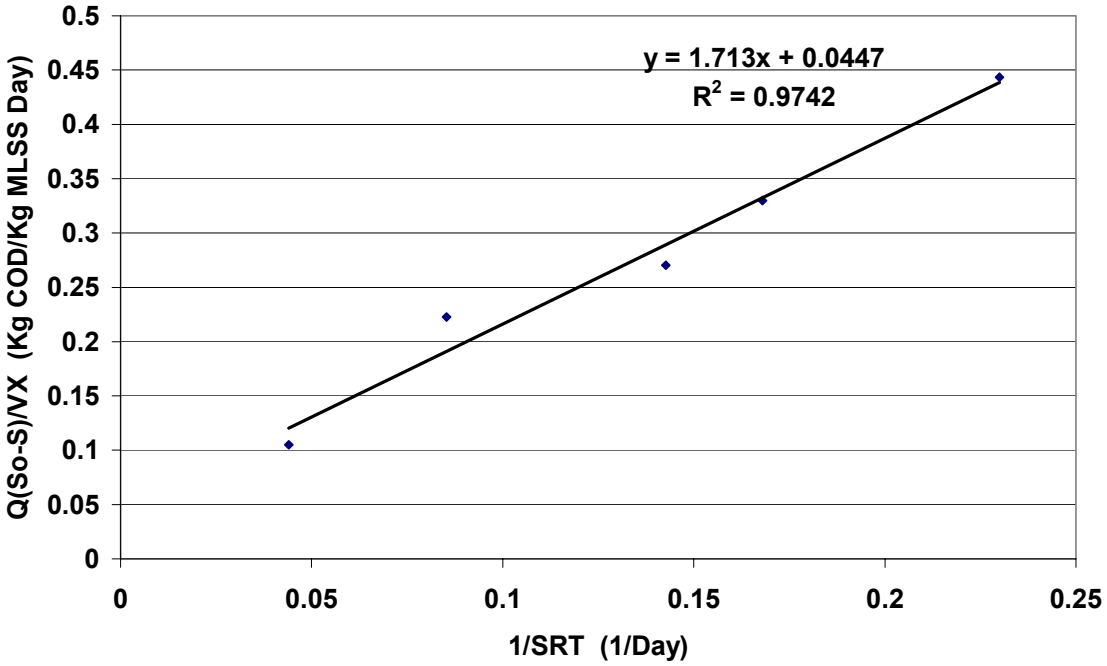

Figure 5.25: Determination of Y and k_d at MLSS 15000 mg/l

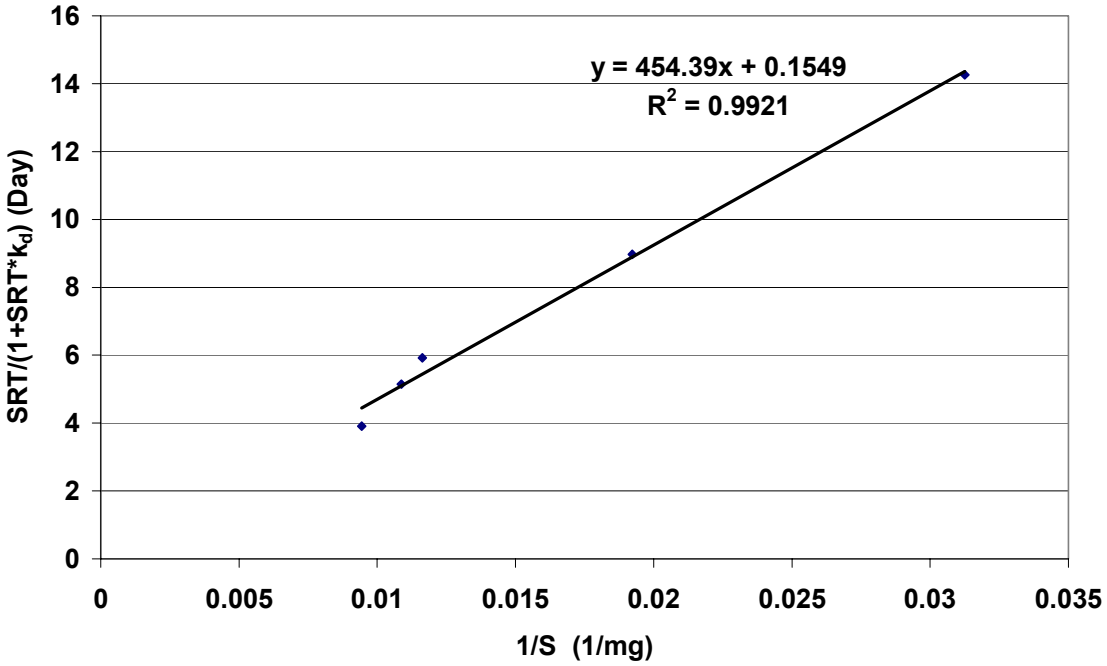

Figure 5.26: Determination of μ_m and K_s at MLSS 15000 mg/l

During this run, the organic loading rate was varied from 0.11 kg COD/ kg MLSS day to 0.45 kg COD/ kg MLSS day. The corresponding highest influent COD concentration applied was 18 450 mg/l COD. Table 5.4 shows the steady state data for the biomass concentration of 15 000 mg/l. Plots of the steady state points, for the determination of the growth kinetics are shown in Figures 5.25 and 5.26. The values of the kinetic coefficients for the 15 000 mg/l of mixed liquor suspended solids were: Y = 0.583 mg/mg, k_d = 0.0261 day^{-1}, μ_m = 6.460 day^{-1} and K_s = 2933 mg COD/l.

5.5.3 Discussion on the determination of the kinetic coefficients

A summary of the kinetic coefficients obtained from the continuous system studies is presented in Table 5.5. It appears that the kinetic coefficients vary significantly with the change in MLSS concentration. However, this variability does not follow any definite pattern, and is not straight forward to draw a firm conclusion. This variability might be attributed to the nature of the system itself, since the system could be a selective process and the kinetic coefficients obtained might represent a different species (Grady and Lim, 1980). This view is supported by an investigation of the performance of the SM-AS unit during the full period of study. For example, looking at the period when the reactor was run at an MLSS 15 000 mg/l, during the first three steady state periods (204 – 229th day), the response of the culture to the increase in organic loading rate appeared to be giving a decrease in COD removal efficiency. But during the next two subsequent steady state periods, COD removal efficiency increased in spite of increase in OLR. This could be attributed to one or two reasons or a combination of both:

Table 5.5: Monod Kinetic Coefficients for an SM-AS Process at Different MLSS Concentrations

MLSS, mg/l	Y (mg/mg)	kd (day-1)	μm, (day-1)	KS(mgCOD/l)
3000	0.487	0.151	1.28	289
5000	0.567	0.062	1.398	326.14
10000	0.571	0.037	5.52	1967
15000	0.583	0.0261	6.46	2933

i. Since the growth rate was controlled by the SRT which was carried out daily by wasting a certain volume of mixed liquor, this might have affected the growth kinetics of the microbial populations in the system. Chiu et al (1972) have described the continuous culture process as a competitive process, which results in the enrichment of a bacterial species at a particular SRT, that is species with higher values of specific growth rate (μ) appeared to be predominant at lower SRT while those species having lower value of μ were enriched in the system only at high SRT

ii. At a given MLSS concentration, the system might have an optimum organic loading rate which gives the maximum COD removal efficiency. Because at higher MLSS concentrations, the OLR in no longer linearly proportional to the substrate removal efficiency.

Generally, the values of the kinetic coefficients which are presented in Table 5.5 are within the normal range for the activated sludge process, but the values of K_s, specially for MLSS of 15 000 mg/l are much higher than those reported in the literature. The estimation of the K_s value is affected by an estimation of the decay rate, k_d, thus any uncertainty in estimating k_d will be reflected in K_s.

Table 5.6 summarizes some of the kinetic coefficients obtained from different sources. Other values for the kinetic parameters can be found in Benefield and Randall (1980) and Grady and Lim (1980). Although, Y, k_d and μ_m are within the reported values for the conventional activated sludge process, they also differ quite significantly. The Y values were increasing with increase in MLSS concentrations, since they represent all the

Table 5.6: Kinetic Coefficients Obtained from Different Sources

Substrate	Y (mg/mg)	k_d (day^{-1})	μ_m, (day^{-1})	K_S (mgCOD/l)	Treatment System	Reference
M.W.W	0.48-0.60	0.05-0.16	5.6-8.10	250-3720	CF-ASP	El-Kebir - 1991
Synth.	0.42-0.53	0.05-0.19	0.8-6.3	83-646	ASP	Amiritharajah - 1983
Glucose	0.50-0.62	0.025-0.48	7.4-18.5	11- 181	ASP	Gaudy (1980)
M.W.W	0.4-0.8	0.025-0.075	2-10	15-70	ASP	Metcalf & Eddy - 1991

amount of biomass produced by the growth during the removal of the substrate. Y values were within the range of reported values treating glucose based synthetic wastewaters. However, Y values were lower than that for conventional AS plants treating the municipal wastewaters. This clearly shows that type of substrate has a role to play in the kinetic coefficients.

5.5.4 Simulation of steady state condition

As previously mentioned in section 5.5.2, the derivation of equation 5.17 was based on the assumption that the submerged membrane activated sludge unit was running under steady state conditions. However, to test the validity of equation 5.17 in predicting the permeate COD at various hydraulic retention times, a comparison of simulation results using equation 5.17 and the experimental data was carried out.

The kinetic parameters summarized in Table 5.5 were utilized in the simulation results of the model. Figures 5.27 to 5.29 show the plots of experimental data during steady state conditions and simulation studies at different MLSS concentrations. Plotting all the experimental data as well as the simulated data on the same graph provides an assessment of how well the performance of the unit can be described by the Monod model.

From Figure 5.27, which shows the plot when the reactor was operating at 3000 mg/l MLSS concentration, it can be observed that theoretical permeate COD, concentrations were more than the actual experimental runs. However, the trend of permeate COD is similar in both simulation and experimental runs.

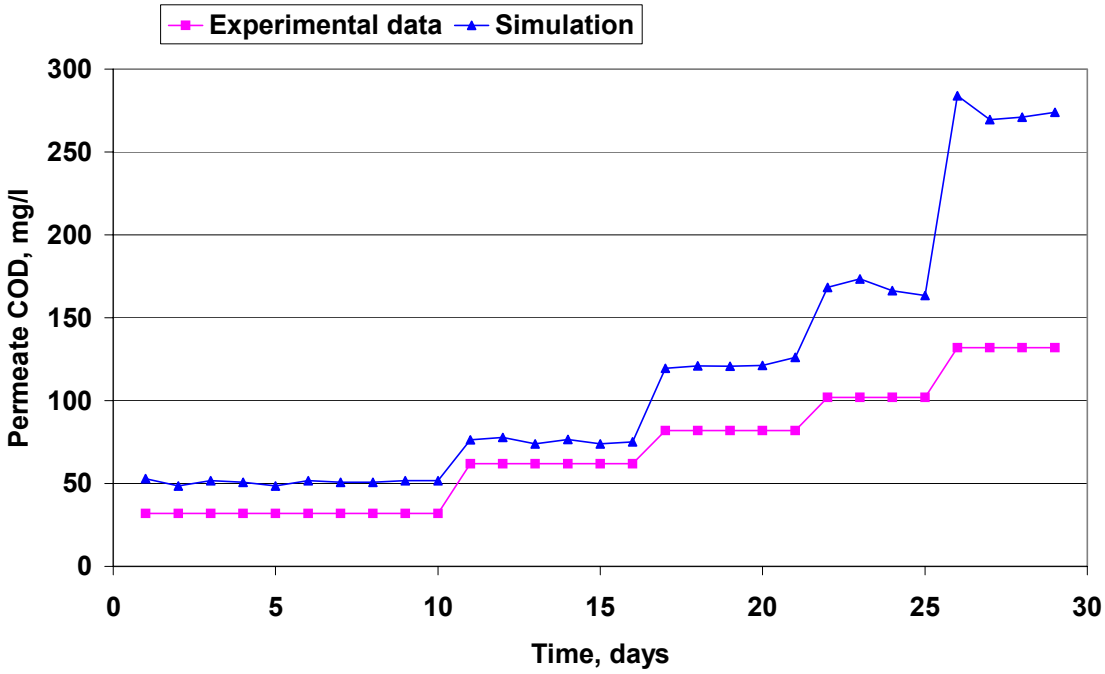

Figure 5.27: Simulation of Effluent COD with Reactor Running at MLSS 3000 mg/l

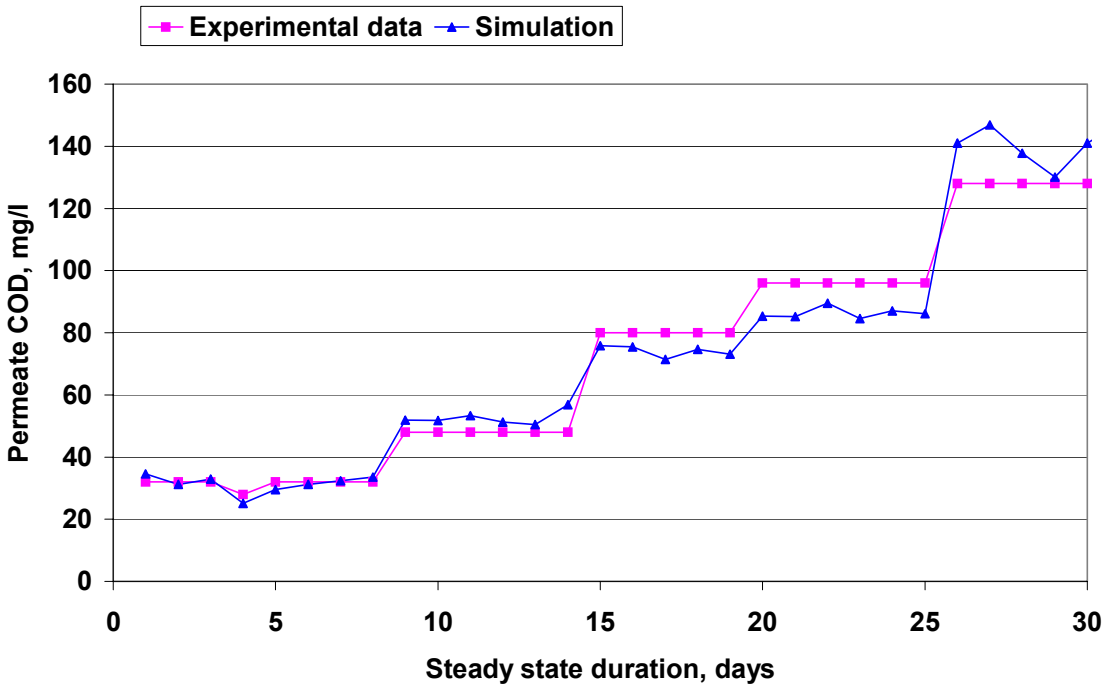

Figure 5.28: Simulation of Effluent COD with Reactor Running at MLSS 5000 mg/l

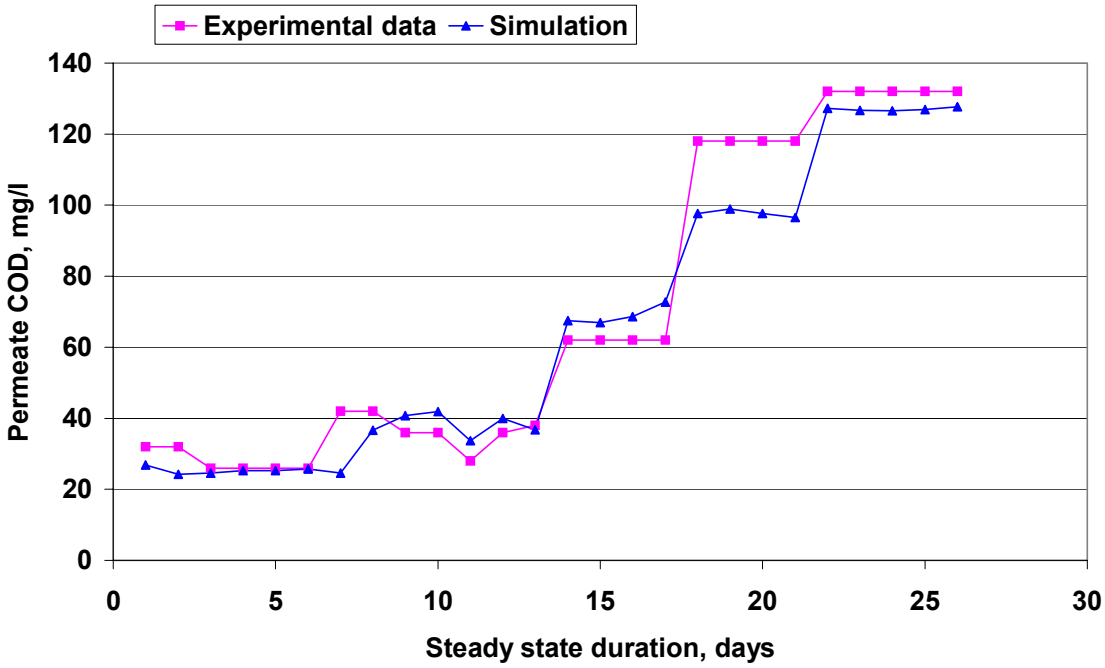

Figure 5.29: Simulation of Effluent COD with Reactor Running at MLSS 10000mg/l

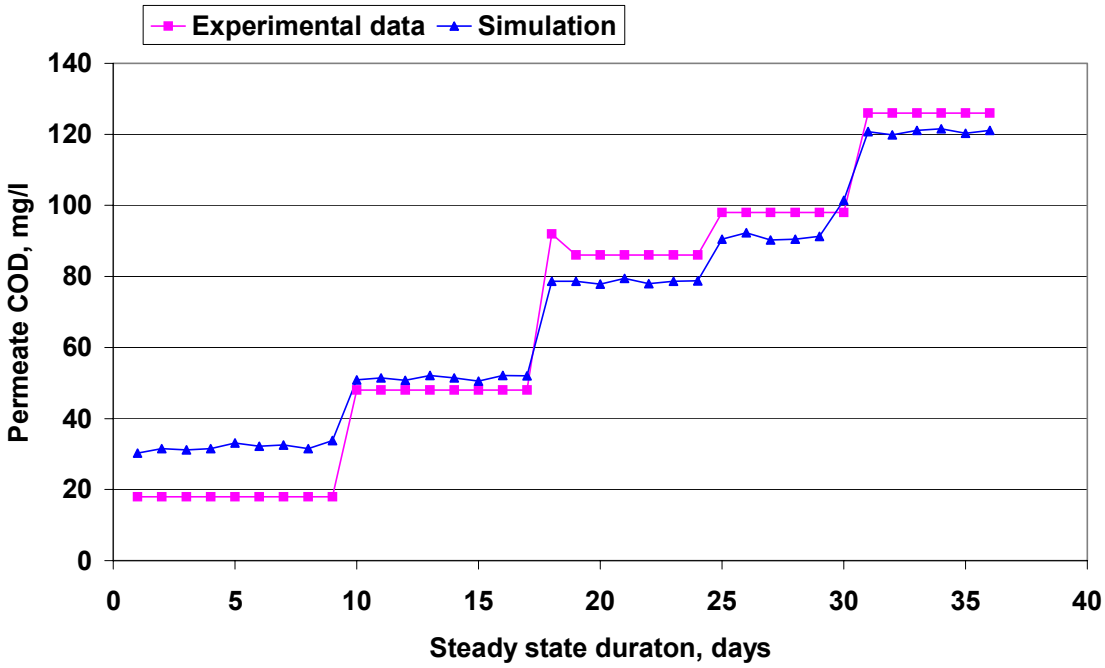

Figure 5.30: Simulation of Effluent COD with Reactor Running at MLSS15000 mg/l

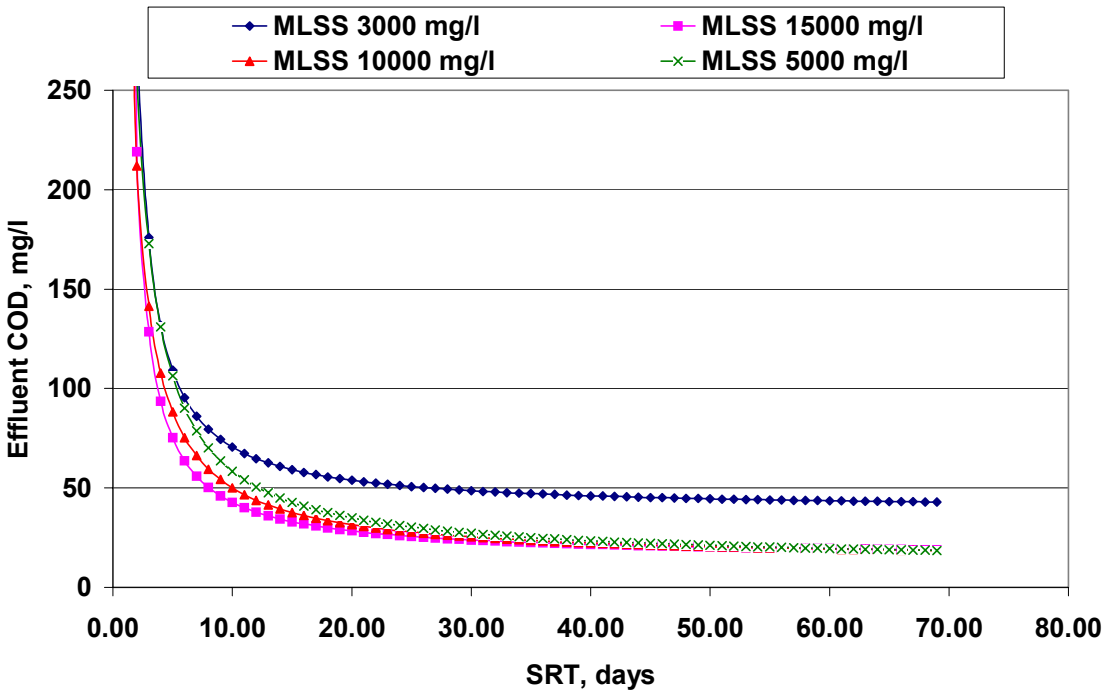

Figure 5.31: Comparison of simulated effluent COD's at various MLSS concentrations

Figures 5.28, shows the permeate COD concentration when the tank MLSS was maintained at 5000 mg/l. Though at lower sludge retention times, the model deviates from the experimental values, Figure 5.28 shows that model fits the steady state experimental data quite well. Figures 5.29 and 5.30 show the simulation results when the reactor was operated at MLSS concentration of 10000 mg/l and 15000 mg/l respectively. These figures show that the model fits perfectly with the experimental data during the steady state period.

Figure 5.31 shows the simulated effluent COD at various MLSS concentrations. This figure clearly shows that up to a certain extent, as the SRT increased, the effluent COD decreased, after this the SRT had no effect on effluent COD concentrations. Also, as the MLSS concentration in the aeration tank increased, the effluent COD decreased, this may be due to decrease in hydraulic loading rate as the tank MLSS concentration increases.

Based on the above discussion it can be concluded that, initially, when MLSS was 3 000 mg/l, the model did not perfectly fit the experimental data, this may be attributed to period of population selection. Initially, a wide variety of species existed, and then progressively selection reduced the range of species. As a result, the range of variation was less obvious, if not present, in the 10 000 mg/l and 15 000 mg/l of MLSS.

5.5.5 Sensitivity Analysis of the Biokinetic Coefficients

In order to determine which biokinetic parameter has the greatest influence on the effluent substrate concentration, a sensitivity analysis was performed. The value of k_d, μ_m and Ks each was varied by ± 50 %, while the other parameters were kept constant. The sludge retention time was kept at 25 days for all these analysis. Sensitivity of the various

biokinetic coefficients were studied by simulating the effluent COD using equation 5.17. The results of the sensitivity analysis are shown in Table 5.7 and in Figures 5.32 to 5.35.

It can be clearly seen that k_d and Ks are directly proportional to the effluent substrate concentration, where as μ_m is inversely proportional to the effluent substrate concentration. Irrespective of tank MLSS concentration, it is observed that effluent substrate concentration was more sensitive to μ_m compared with k_d and Ks. At the MLSS concentration 3000 mg/l and 5000 mg/l, the effluent substrate concentration showed almost equal sensitivity to both k_d and Ks. However at the higher MLSS concentrations, effluent substrate concentration was more sensitive to K_s than to k_d.

From the above sensitivity analysis it is clear that care should be taken when using these biokinetic coefficients in the Monod model for the designing of submerged membrane bioreactors. Extra caution should be exercised when μ_m is dealt with, since small variation in μ_m can result in big change in effluent substrate concentration.

Table 5.7: Variation of Biokinetic Coefficients for Sensitivity Analysis

Tank MLSS mg/l	**Biokinetic Parameter**	**Based on Current Study**	**50 % Decrease**	**50 % Increase**
3000	k_d (day^{-1})	0.151	0.0755	0.2265
	μ_m (day^{-1})	1.28	0.64	1.92
	K_s (mg COD/l)	289	144.5	433.5
5000	k_d (day^{-1})	0.062	0.031	0.093
	μ_m (day^{-1})	1.398	0.699	2.097
	K_s (mg COD/l)	326.14	163.07	489.21
10 000	k_d (day^{-1})	0.037	0.0185	0.0555
	μ_m (day^{-1})	5.52	2.76	8.28
	K_s (mg COD/l)	1967	983.5	2950.5
15 000	k_d (day^{-1})	0.0261	0.01305	0.03915
	μ_m (day^{-1})	6.46	3.23	9.69
	K_s (mg COD/l)	2933	1466.5	4399.5

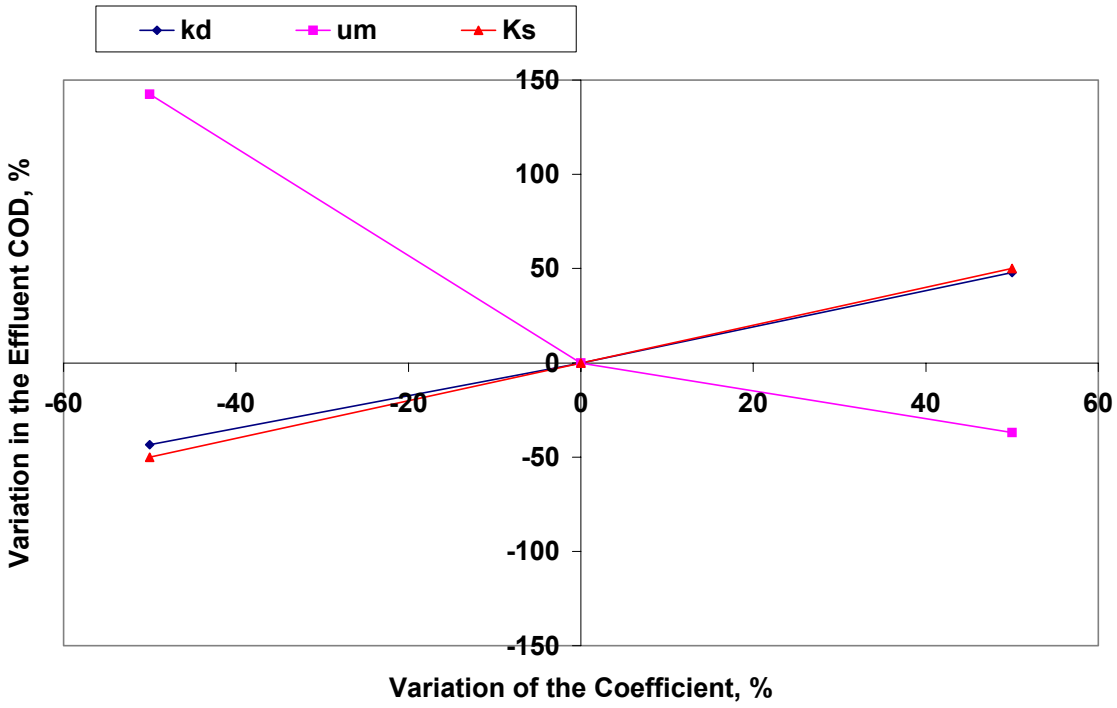

Figure 5.32: Sensitivity of Biokinetic Coefficients at MLSS 3000 mg/l

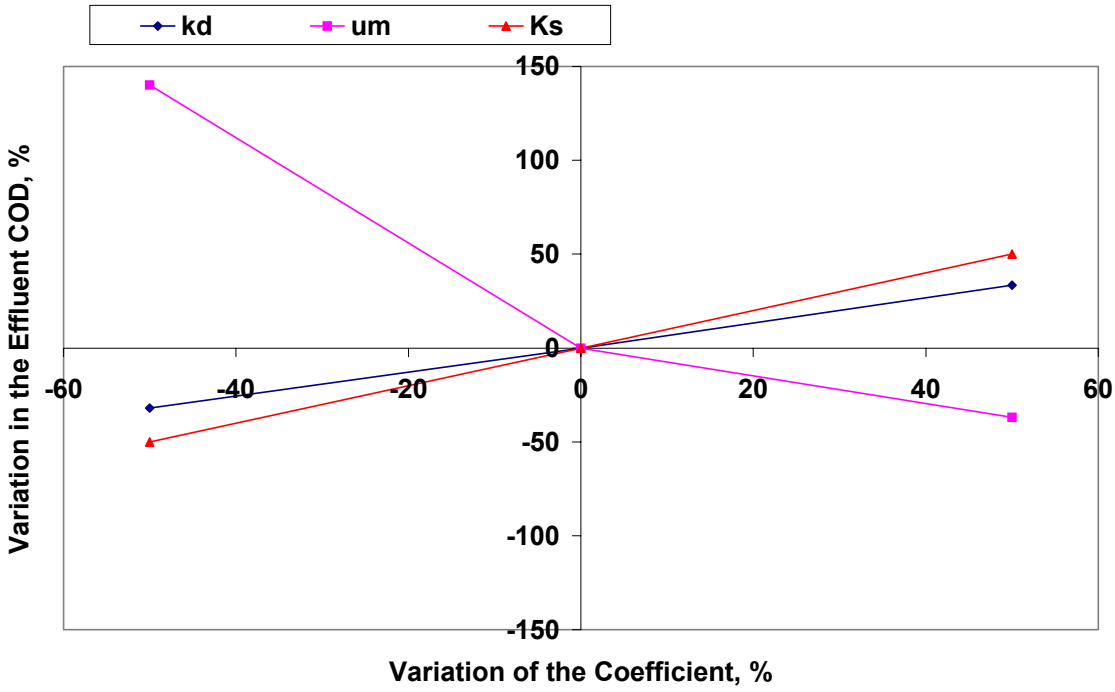

Figure 5.33: Sensitivity of Biokinetic Coefficients at MLSS 5000 mg/l

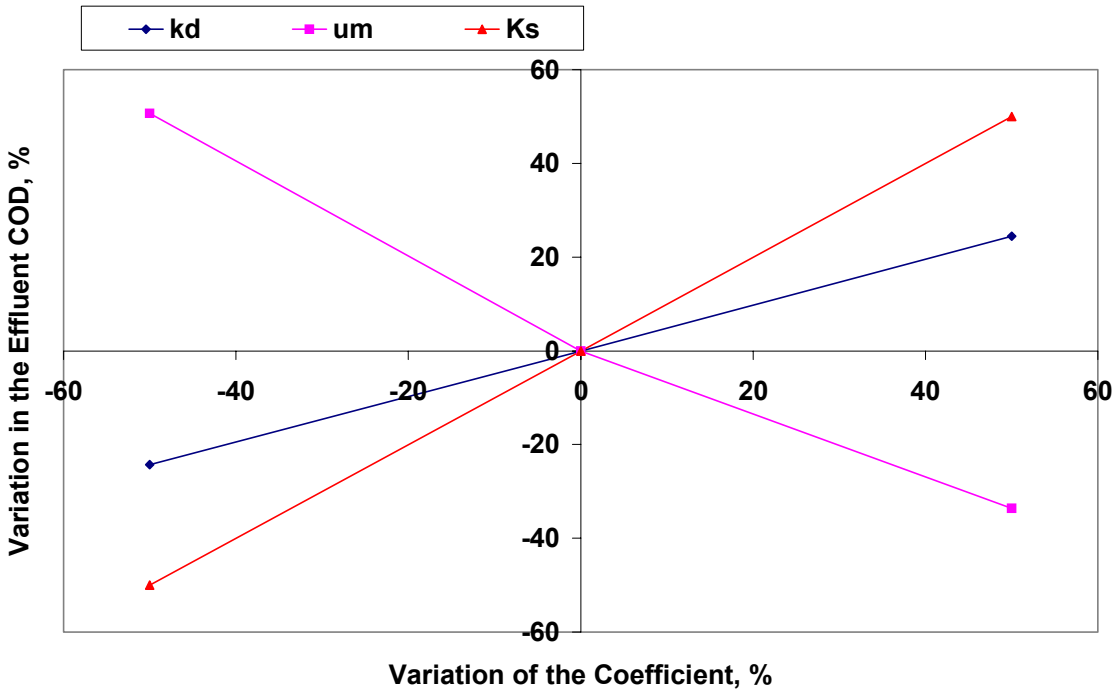

Figure 5.34: Sensitivity of Biokinetic Coefficients at MLSS 10000 mg/l

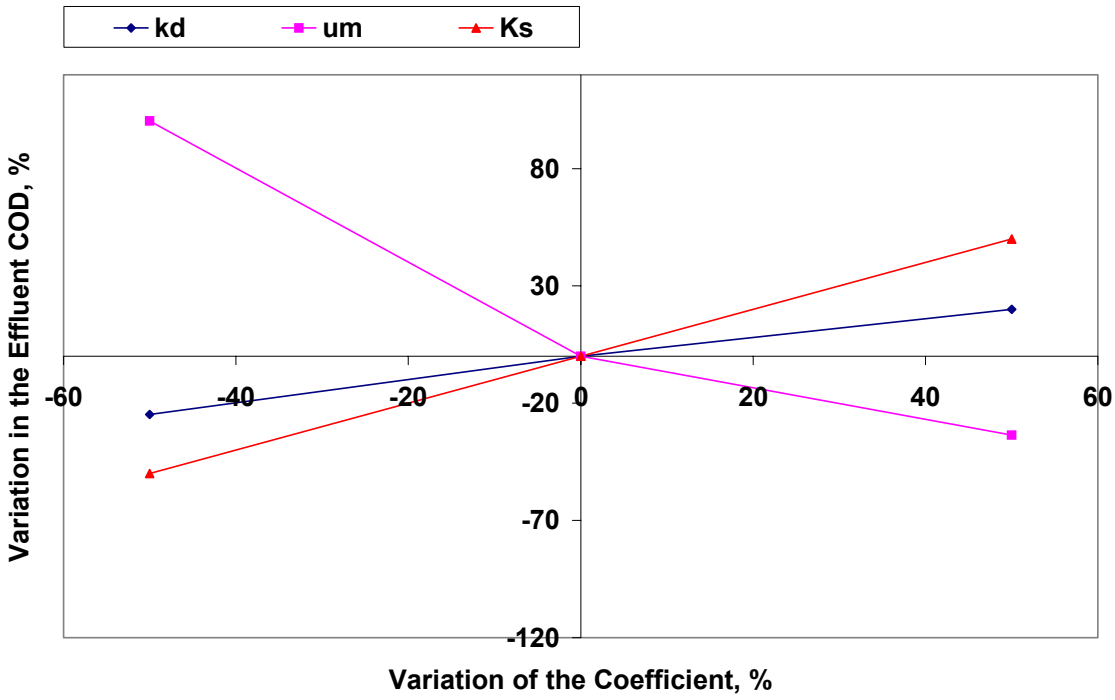

Figure 5.35: Sensitivity of Biokinetic Coefficients at MLSS 15000 mg/l

5.6 MICROBIAL ANALYSIS

This study has two objectives, first to find out the viable bacterial count in the reactor at various reactor mixed liquor concentrations and secondly to identify the microorganisms which survived and adopted the higher MLSS concentrations. For determination of viable bacterial count, heterotrophic plate count method was adopted. Both tank MLSS and permeate were analyzed for the microorganisms.

The pour plates were incubated for 48 hours at 35^{o}C. Following the incubation, it was noticed that there were only two types of colonies that survived: big spongy white colonies and small whitish yellow colonies. Table B1 and Figure 5.36 shows the variation of plate count results of both tank and permeate at different tank MLSS concentrations. As expected the plate count increased with increase in MLSS, showing that viability is maintained at higher MLSS concentrations as well.

It is important to note that the permeate plate count values showed far less numbers than the tank values, this shows that the microfiltraion membrane is effectively filtering the microorganisms. Almost four logs reduction is permeate plaque forming units were observed at the tank MLSS concentration of 15 000 mg/l.

Samples of each type of colonies were isolated and incubated in the same medium at the same temperature for another forty eight hours. This was done to isolate the bacteria for species identification. The identification was carried out at microbiology lab of King Faisal University, Dammam. The API 20E kit was used for carrying out identification. The identification showed that the:

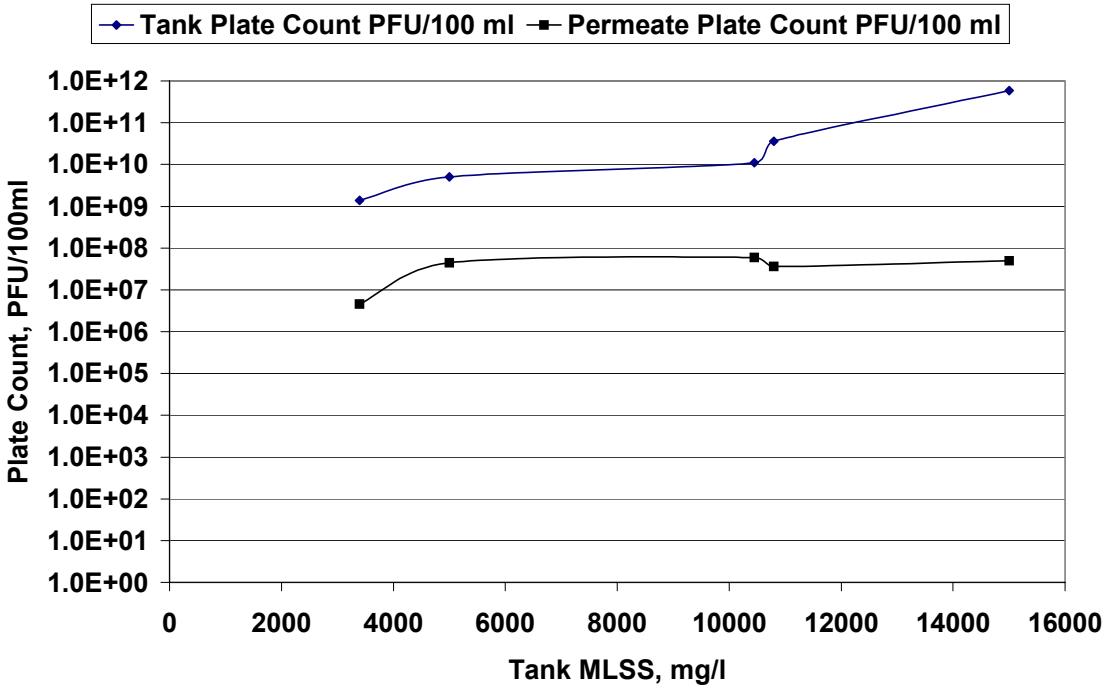

Figure 5.36: Plate Count Results

Big spongy white colonies were *Serratia liquefaciens* and

Small whitish yellow colonies were *Aeromonas hydrophila*

Extent of removal of indicator bacteria by the microfiltration membrane was also studied. Table B2 to B4 and Figures 5.37 to 5.39 show variation of tank and permeate total coliform, feacal coliform and feacal streptococci numbers respectively at different tank biomass concentrations. The figures show that most probable number (MPN) values of these indicator organisms increased both in the tank and permeates with the increase in MLSS concentrations. Especially this increase was more evident at MLSS 15000 mg/l. It is also evident that the concentration of these indicator microorganisms in the permeate were much less than in the tank indicating these organisms were retained by the microfiltration membrane.

Though in the literature, permeate Feacal coliform and Feacal streptococci were reported as low as one and below detectable limits respectively (Ueda and Horan, 2000), the same efficiency could not be achieved with the polyester woven fabric microfiltration membrane. This could be due to larger size of the membrane pores.

Another significant observation that can be made from these figures is that the removal of indicator microorganisms was (for example, seven logs in case of Feacal streptococci at MLSS 15000 mg/l) higher at 15000 mg/l MLSS concentrations. This could be due to pore clogging due to accumulation of the end products.

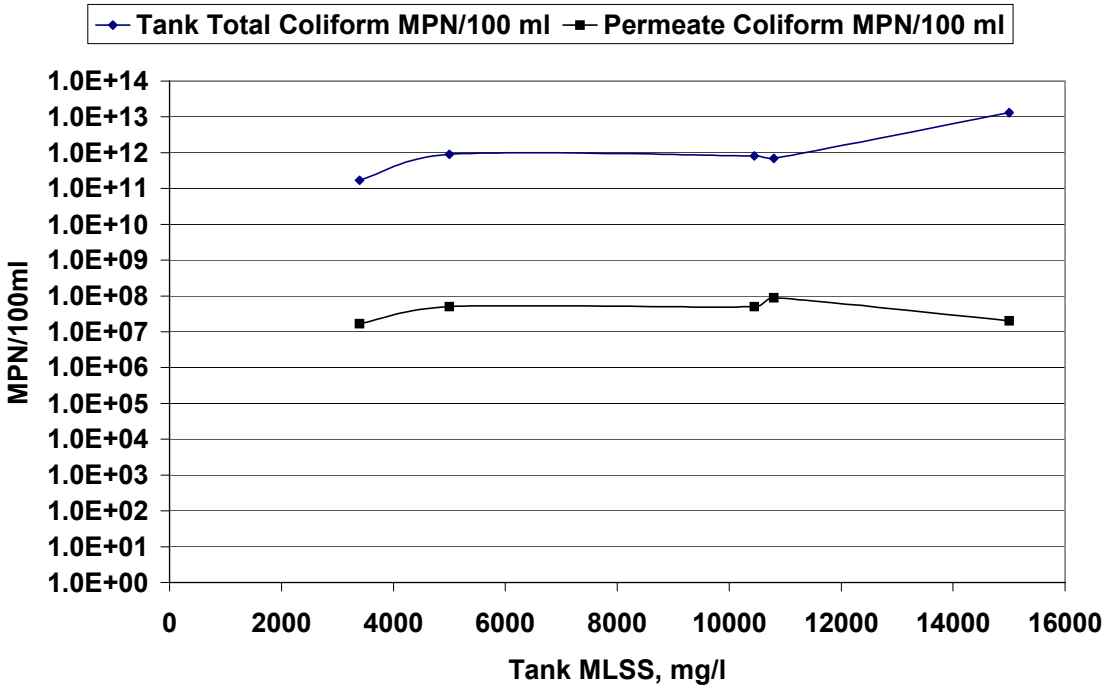

Figure 5.37: Variation of Total Coliform

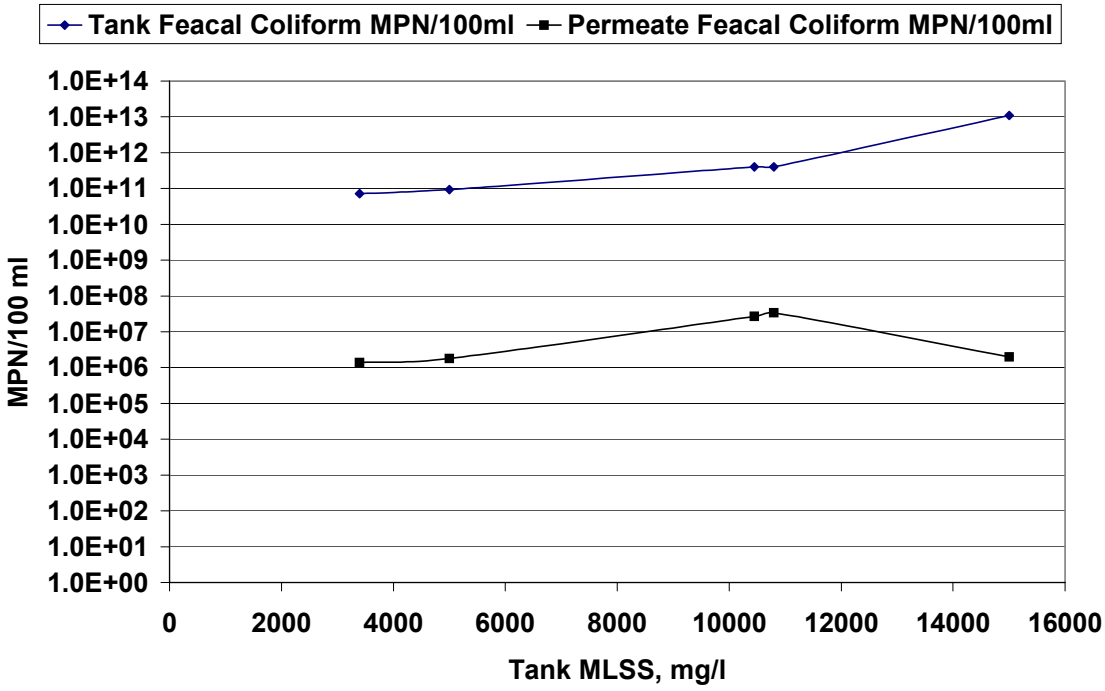

Figure 5.38: Variation of Feacal Coliform

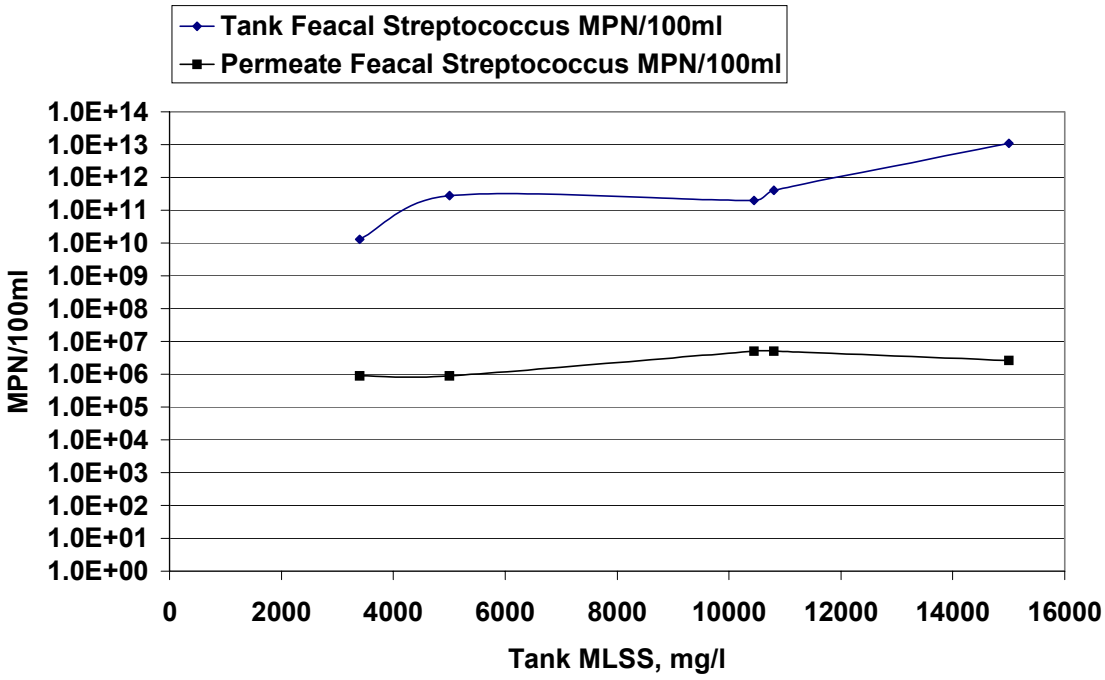

Figure 5.39: Variation of Feacal Streptococci

5.7 SHOCK LOADING - SUDDEN INCREASE IN OLR

Applying the shock loadings assessed the response and stability of the submerged membrane bioreactor to a sudden increase in organic loading and subsequent degree of disruption of substrate removal. Two alternatives might be applied to test the system response, hydraulic shock loading or organic shock loading. The former was not adopted due to the limitation imposed by the flux of the membrane, and an organic shock load was applied by changing the influent COD concentration.

Three trials of shock loadings were carried out when the tank MLSS concentration was at 15000 mg/l. In the first trial on the 252nd day, the OLR was increased from 0.41 kg COD/ kg MLSS day to 0.72 kg COD/ kg MLSS day. This change increased the influent COD concentration from 4998 mg/l to 8024 mg/l as shown in Figure 5.40. This loading was maintained for two days, then on 254th day OLR was increased to 1.03 kg COD/ kg MLSS day, corresponding to influent COD of 12100 mg/l. The tank MLSS and permeate quality was monitored. This loading was maintained for another day and on 256th day the organic loading rate was further increased to 1.22 kg COD/ kg MLSS day corresponding to 16 000 mg/l influent COD.

Figure 5.41 and 5.42 shows the permeate COD concentration and COD removal efficiency, respectively, during the shock loading period (Table C1). Though influent substrate concentrations doubled and tripled, the permeate COD concentrations did not reflect these shock loadings as evident from the Figure 5.41. This shows that the SM-AS system at an MLSS concentration of 15 000 mg/l could easily with stand these shock loadings. Figure 5.43 shows the variation of OLR and SLR during the shock loading

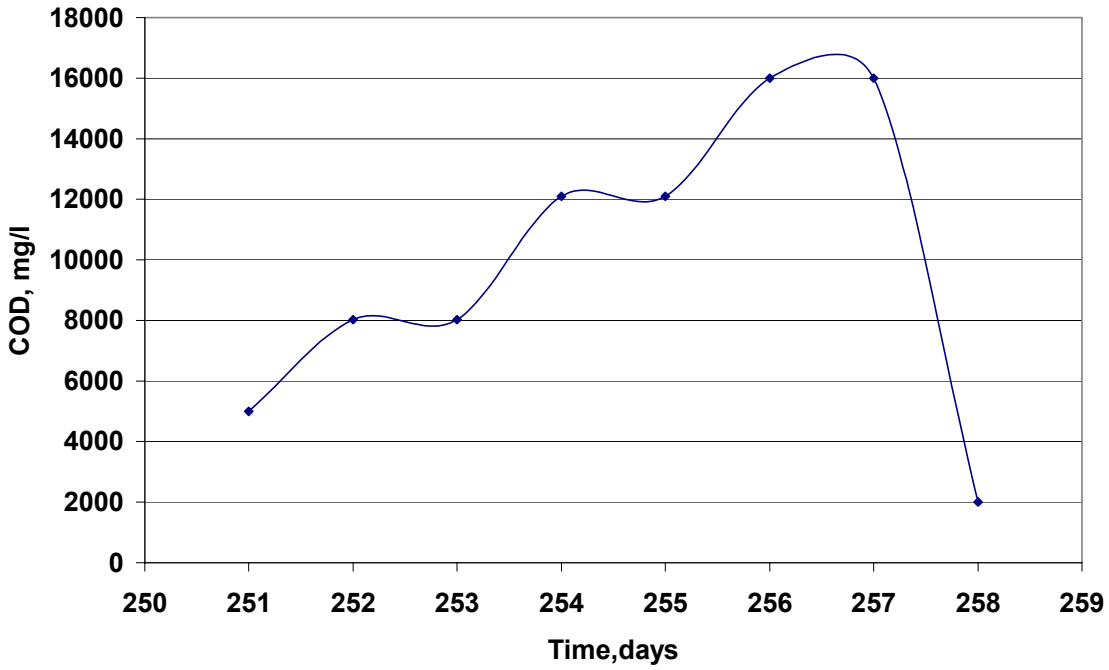

Figure 5.40: Influent Substrate Concentration for Shock Loading Runs

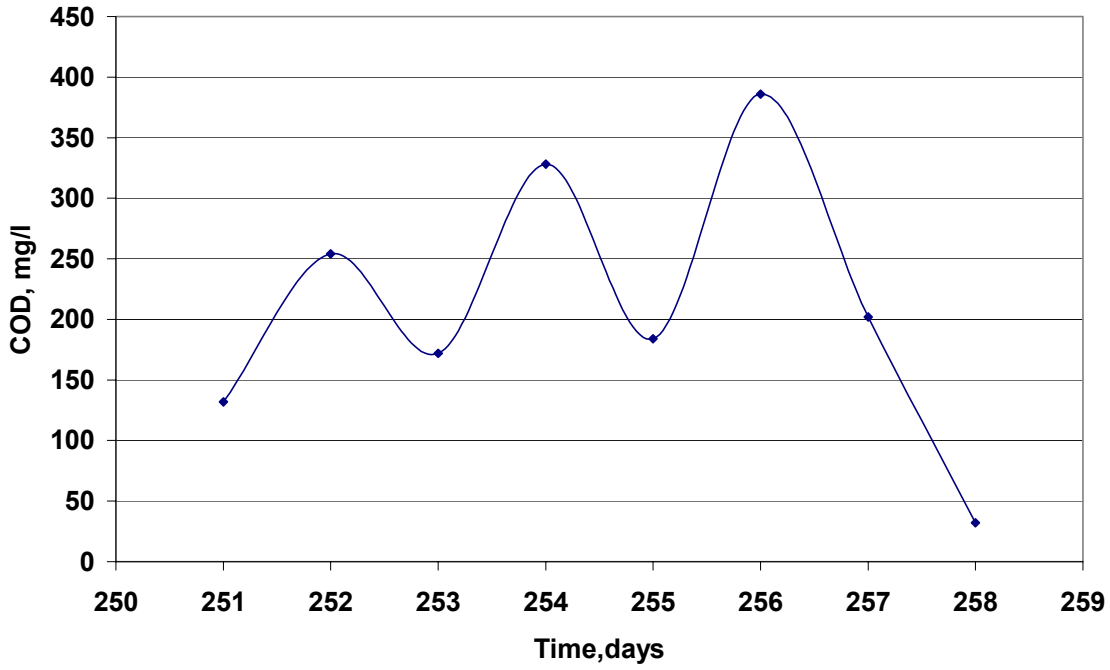

Figure 5.41: Permeate COD for Shock Loading Runs

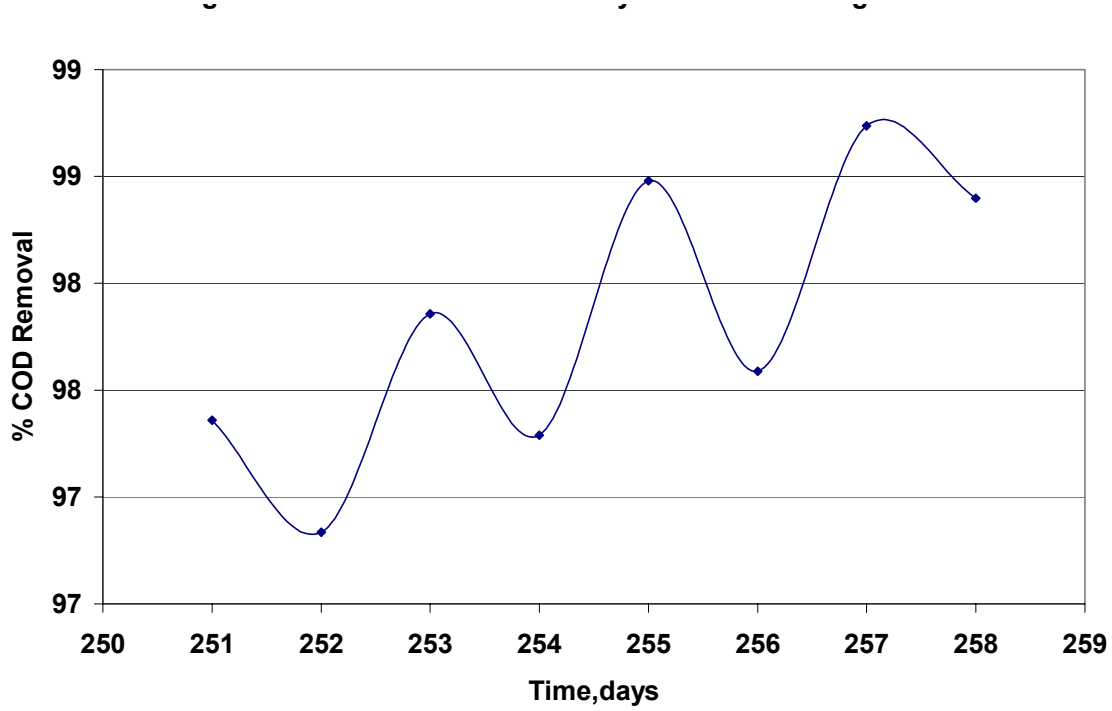

Figure 5.42: COD Removal Efficiency for Shock Loading Runs

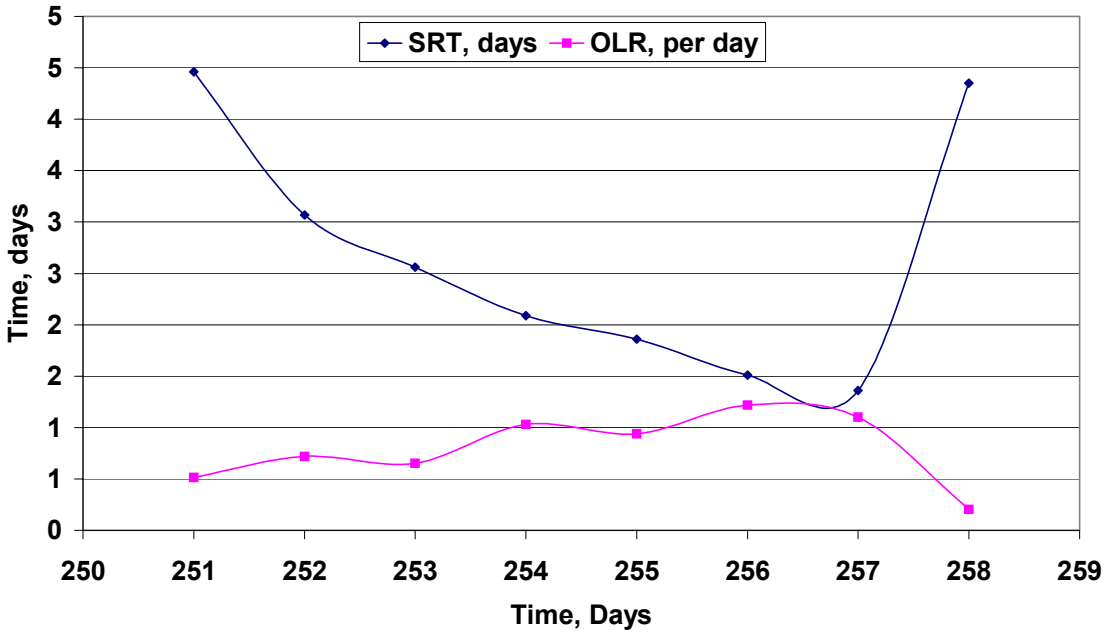

Figure 5.43: Variation of SRT and OLR for Shock Loading Trials

trials. It can be observed that as the OLR increased, the SRT decreased. This was due to more sludge production at higher organic loading rates. The MLSS was kept constant at 15000 mg/l during these studies by wasting the excess sludge once a day. On the 258th day, when the loading was returned back to normal loads (0.2 kg COD/ kg MLSS day), the permeate COD returned back to normal value (32mg/l COD).

5.8 SHOCK LOADING -TOXIC LOADING

The ability of the system to withstand toxic loadings and extent influence on substrate removal efficiency was studied by adding two different types of toxicants. One is organic toxicant, phenol and the other is metallic toxicant chromium.

5.8.1 Phenol

The concentration of phenol in the industrial wastewaters ranges from 100 mg/l to 1000 mg/l (Al-Kassim et al. 1993; Wu et al. 1993). It is important to remove phenols and aromatic compounds from contaminated industrial wastewaters before discharge into any water body because of their toxicity to aquatic organisms. Conventional treatment methods such as chemical, physical and biological processes are not always suitable for treating moderate to high concentration wastewaters (Villalobos and Buchanan, 2002). The microorganisms in the activated sludge process are susceptible to chemical substances such as phenol (Blum and Speece, 1991).

The study was carried out both before and after acclimatizing the microorganisms to phenol. A toxic shock loading of 400 mg/l of phenol was added to the synthetic substrate of COD 992 mg/l. Figure 5.44 shows the performance of the SM-AS unit for two weeks

under this shock loading (Table C2). The removal efficiencies of both COD and Phenol dropped drastically to a low of 9 % and 14 % respectively. With the continuation of the phenol in the effluent, the removal efficiencies gradually improved and reached steady state values of 77 % for COD and 70% for phenolic removal efficiency by thirteenth day. The bacteria might have acclimatized to the phenolic feed thus giving steady phenolic and COD removal efficiencies.

Yamagishi et al. 2001, were able to completely remove phenol in a Cross-flow activated sludge process, with a tank MLSS concentration of 8000 mg/l. But the bacteria were acclimatized to the phenolic feed for over 15 years. Therefore, Performance of the SM-AS unit to phenolic loadings after acclimatizing the bacteria to phenol for over a month was also investigated.

Initially, phenolic concentration of 50 mg/l was added to the substrate COD of 998 mg/l. After fifteen day, the phenolic concentration was increased to 100 mg/l for another two weeks. The influent concentration was kept constant at 998 mg/l during entire toxic loading study period. Then 316^{th} day, the influent phenol concentration was increased to 400 mg/l for the next five days. After that, the influent phenol concentration was increased to 800 mg/l for the next four days.

Figure 5.45 shows the variation of effluent substrate, influent phenol and effluent phenol during the toxic loading duration (Table C3). Figure 5.46 shows the COD and phenol removal efficiencies during the study period. It is evident from these figures that even at 400 mg/l phenol; COD removal of over 80% could be achieved. Also over 75% of phenol was removed by the SM-AS system. At influent phenolic concentration of 800 mg/l, the.

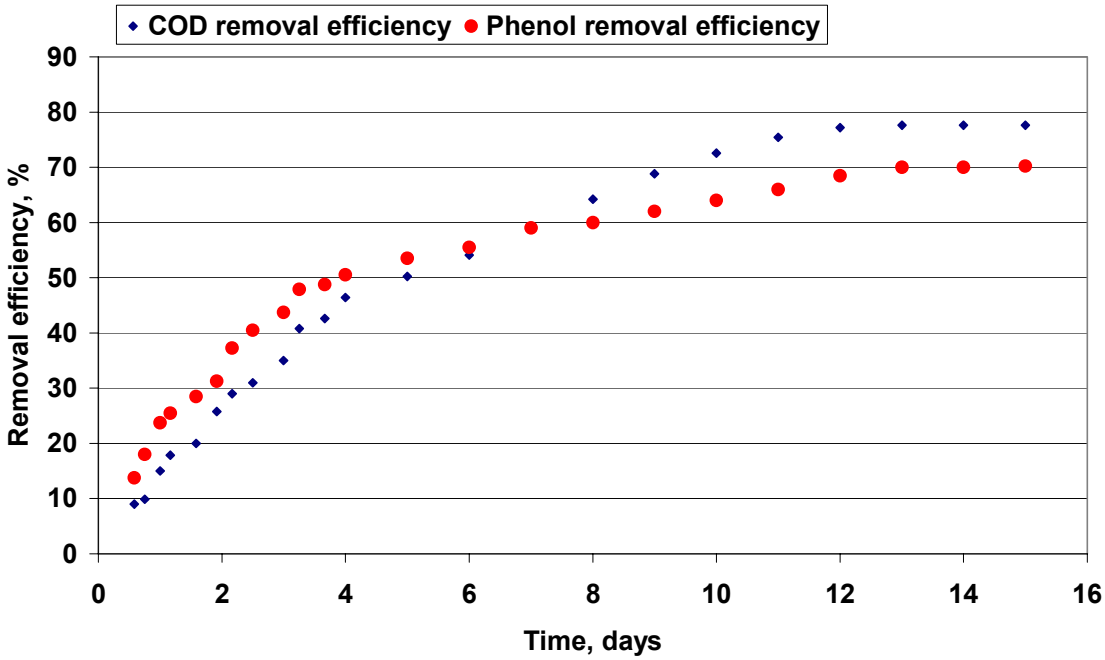

Figure 5.44: Performance of the SM-AS unit under shock loading of 400 mg/l Phenol

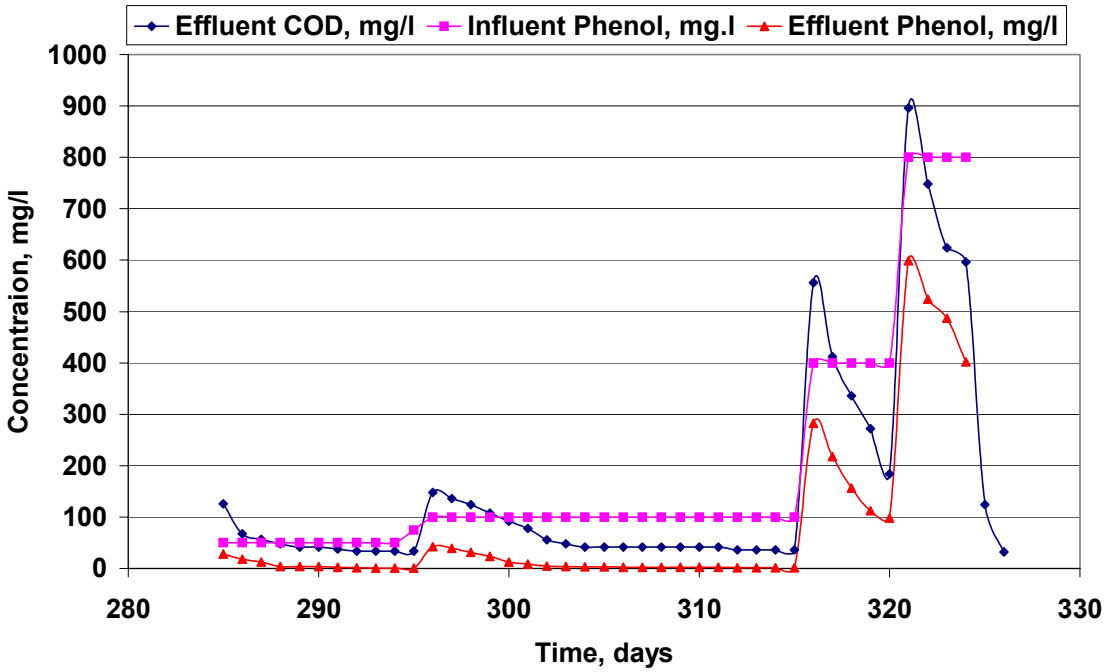

Figure 5.45: Effect of Phenol Loading (with acclimatization)

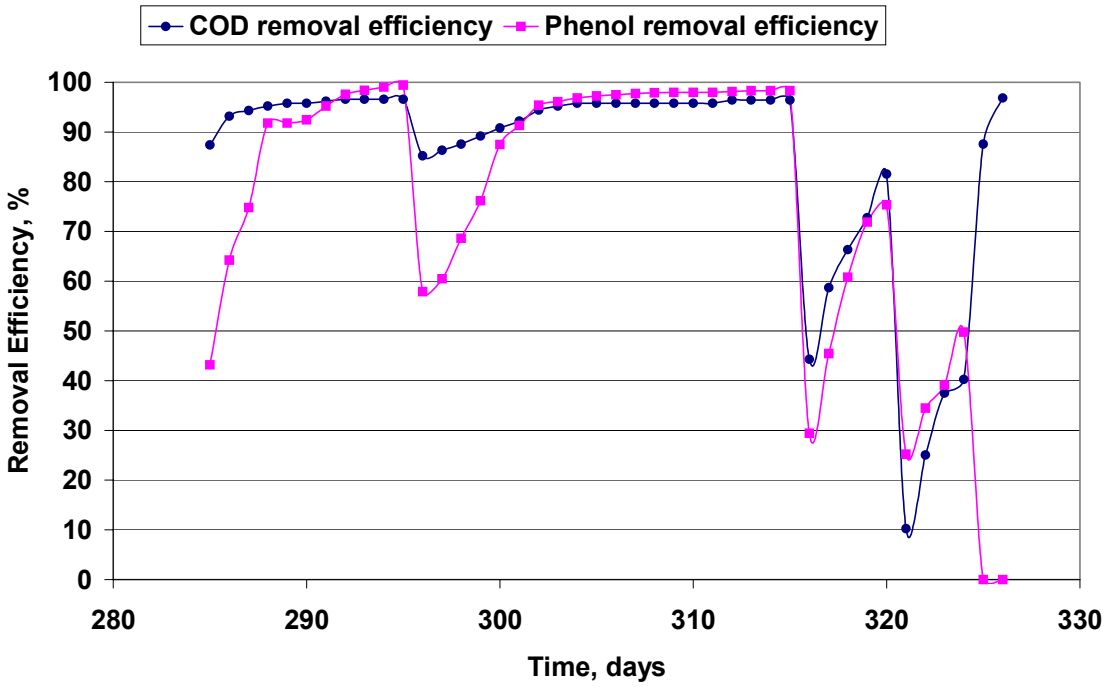

Figure 5.46 Removal Efficiencies under Phenolic Loading (With acclimatization)

COD removal efficiency dropped to less than 40 % indicating that phenol is inhibiting the bacterial action in the aeration tank.

At this point, phenolic load was removed, and surprisingly the reactor bounced back to its original treatment capacity (effluent COD of 28 mg/l). This shows that the SM-AS system at 15 000 mg/l could withstand a high phenol loading of 800 mg/l without adversely affecting the system and the systems would bounce back to its effluent quality within a day after removal of the toxic loading

5.8.2 Chromium loading

Chromium concentrations greater than 11 mg/l, reduces the maximum growth rate in a conventional activated sludge process. Generally 5 mg/l chromium concentration is the threshold limit allowed in most of the conventional AS processes (Nemerow, 1978). To study the effect of this metallic toxicant on the substrate removal efficiencies in a SM-AS process, three concentrations of chromium were applied. Initially a 20 mg/l chromium was supplied with 998 mg/l influent COD. Then on the third day, the chromium dosage was doubled to 40 mg/l. After another day, the chromium concentration was increased to 50 mg/l. At this stage the permeate COD removal efficiency dropped to 26 %.

Table C4 and Figure 5.47 shows the variation of permeate COD, influent and effluent chromium concentrations respectively. Figure 5.48 shows the COD and chromium removal efficiencies. It is seen that, a chromium concentration of 50 mg/l adversely affects the performance of the SM-AS process. However, when the toxic loading was removed, the system bounced back to its original treatment efficiency within two days.

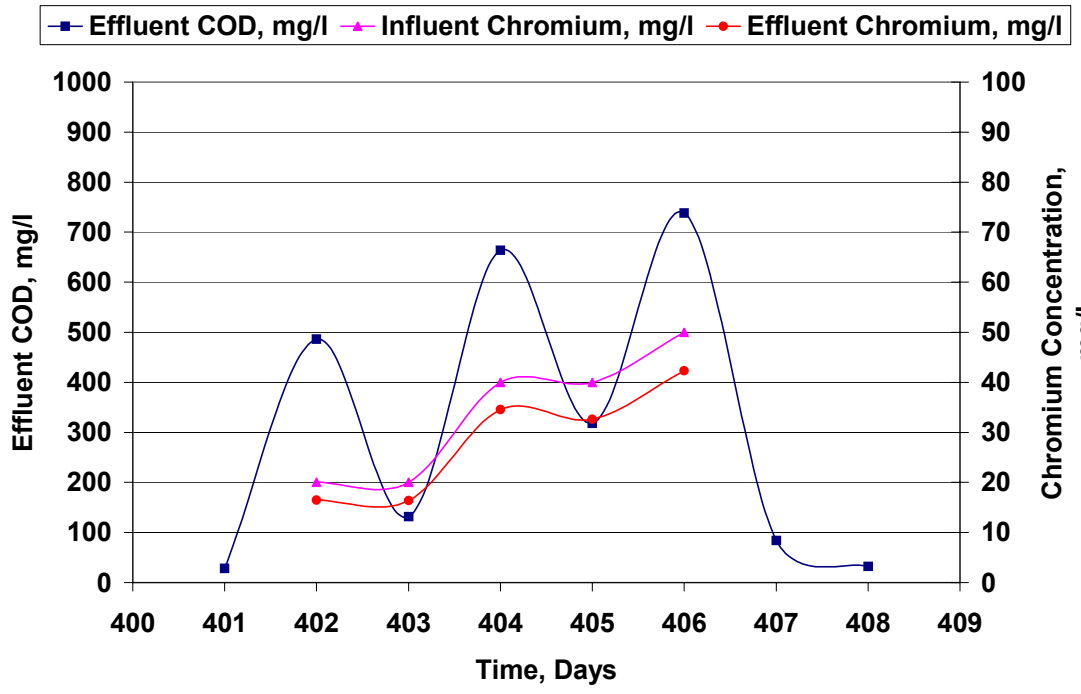

Figure 5.47: Effect of Chromium Loading on the SM-AS System

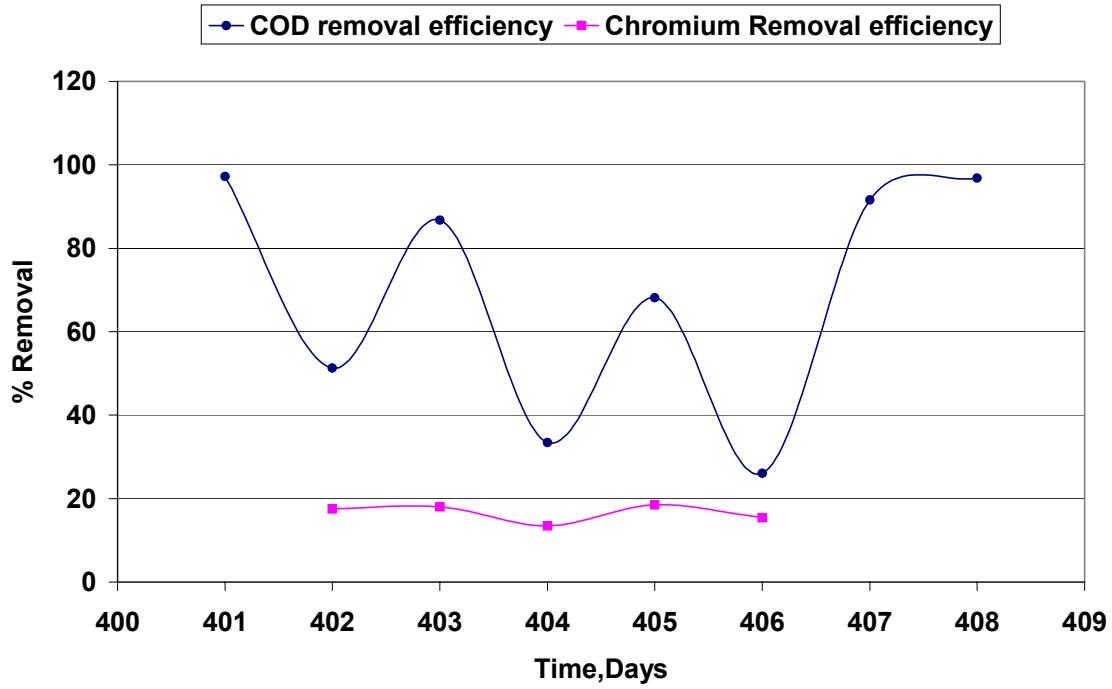

Figure 5.48: Removal Efficiencies Under Chromium Loading

CHAPTER 6

CONCLUSIONS

The following conclusions are made from the present research:

- Treatment efficiency of the SM-AS process under different operating conditions (such as MLSS, OLR and SRT) was studied. The COD removal efficiency ranged from 80% to over 98%. The COD removal efficiency increased as the MLSS concentration increased.
- At MLSS concentration of 3000 mg/l, the kinetic coefficients were: Y = 0.487 mg/mg, k_d = 0.151 day-1, μ_m = 1.28 day^{-1} and K_s = 289 mg COD/l. An organic loading rate of 1.07 kg COD/ kg MLSS day gave the maximum COD removal efficiency (95.6% COD removal) at 3000 mg/l MLSS concentration.
- At MLSS concentration of 5000 mg/l, the values of the kinetic coefficients were: Y = 0.567 mg/mg, k_d = 0.062 day^{-1}, μ_m = 1.398 day^{-1} and K_s = 326 mg COD/l. During this runs, variation of OLR did not had much influence on the COD removal efficiency. The COD removal efficiencies were in the narrow range of 94 – 95%COD removal.

- The values of the kinetic coefficients for the 10 000 mg/l of mixed liquor suspended solids were: Y = 0.571 mg/mg, k_d = 0.037 day^{-1}, μ_m = 5.520 day^{-1} and K_s = 1967 mg COD/l. Maximum COD removal efficiency (97%) was achieved at an OLR of 0.32 kg COD/ kg MLSS day.
- The values of the kinetic coefficients for the 15 000 mg/l of mixed liquor suspended solids were: Y = 0.583 mg/mg, kd = 0.0261 day-1, μm = 6.460 day-1 and Ks = 2933 mg COD/l.
- The model based on Monod kinetics fits perfectly with the experimental data during the steady state period for MLSS concentrations 10000 mg/l and 15000 mg/l.
- The microorganisms involved are identified as Big spongy white colonies of Serratia liquefaciens and Small whitish yellow colonies of Aeromonas hydrophila
- Fouling of the membrane was controlled by intermittent pumping schedule, back flushing with air and/or mechanical cleaning with brush.
- The system could withstand easily shock loading of 16000 mg/l COD @ an MLSS concentration of 15000 mg/l giving over 98 % COD removal efficiency.
- The system could withstand phenolic concentrations as high as 400 mg/l, giving an effluent COD removal of 77% without acclimatization and COD removal of 81% with acclimatization.
- The toxic loading (after acclimatization) of 800 mg/l of phenol didn't have adverse affect on the system, and it regained its treatment efficiency, within a day after toxic load withdrawal

- The system could withstand easily chromium concentration of 20mg/l, to give an 86% COD removal
- The shock loading of 50 mg/l of chromium didn't have adverse affect on the system, and it regained its treatment efficiency, within two days after toxic load withdrawal.
- The ability of the SM-AS system to withstand shock loading and toxic loading shows that this system can be used for treatment of combined municipal and industrial wasters.

CHAPTER 7

RECOMMENDATIONS

The following recommendations are made for future study in this area:

- Because of the coarse nature of the membrane, the COD removal observed was due to biological oxidation in the reactor. Other membrane material should be tested to see if membrane is also contributing to the COD removal,
- The experiments were performed with a coarse MF membrane. Another type of MF or UF membrane should also be tested for better comparison of the results,
- The lab scale experiments should be tested in the field with bench scale set up, treating real municipal wastewater,
- Studies on retrofitting of existing wastewater treatment plants with SM-AS process should be seriously investigated,
- The ability of the system to withstand shock loading and toxic loading with respect to removal of other parameters like ammonia and phosphates needs to investigated,
- Only phenol and chromium were tested for toxic loading effects, other toxic organic and inorganic chemicals should be tested for toxic loading tests,

- As the results of toxic loading show resilience of SM-AS process for toxic pollutants is high, the scope of applying SM-AS systems for treating combined wastewaters need to be investigated.

APPENDIX A

RAW DATA FOR BIOKINETIC STUDIES

Table A1: Raw Data for Biokinetic Studies

TIME, Days	INFLUENT COD, mg/l	EFFLUENT COD, mg/l	PERMEATE VOLUME, l	TANK MLSS, mg/l		TANK MLVSS, mg/l
				Before Wasting	After Wasting	
1	498	96	40	3060	2980	2560
2	498	96	40	3090	2970	2640
3	498	80	38	3050	2950	2680
4	498	96	38	3020	3020	2670
5	498	80	38	3100	2980	2670
6	498	80	36	3090	3000	2710
7	498	96	36	3100	2980	2700
8	498	80	35	3040	2980	2540
9	498	48	35	3050	2990	2510
10	498	48	35	3030	2990	2560
11	498	48	40	3070	2950	2620
12	498	48	39	3060	2970	2570
13	498	32	39	3040	3040	2590
14	498	32	38	3110	3000	2640
15	498	32	38	3080	3000	2650
16	498	32	38	3080	2990	2610
17	498	32	40	3110	2980	2650
18	498	32	40	3060	3000	2570
19	498	32	39	3090	2990	2610
20	562	40	38	3100	2990	2580
21	562	40	40	3100	3000	2690
22	562	32	40	3120	2980	2730
23	562	32	40	3100	3000	2670
24	562	32	39	3120	2990	2660
25	562	32	39	3100	2980	2700
26	562	32	38	3110	3010	2710
27	562	32	40	3130	3000	2710
28	562	32	40	3120	3000	2700
29	562	32	39	3130	3010	2730
30	562	32	40	3130	3000	2690
31	562	32	40	3120	2990	2680
32	697	64	40	3110	3000	2680
33	697	80	39	3130	3020	2750
34	697	42	22	3030	3030	2450
35	697	48	38	3180	3000	2650
36	697	48	38	3160	3020	2760

Table A1: Raw Data for Biokinetic Studies (Cont.)

TIME, Days	INFLUENT COD, mg/l	EFFLUENT COD, mg/l	PERMEATE VOLUME, l	TANK MLSS, mg/l		TANK MLVSS, mg/l
				Before Wasting	After Wasting	
37	697	48	37	3200	2980	2800
38	697	48	36	3190	2960	2770
39	697	48	36	3200	3000	2720
40	697	42	20	2980	2980	2340
41	996	160	35	3210	3000	2720
42	996	128	40	3280	2990	2840
43	996	96	39	3300	3000	2910
44	996	78	38	3340	3010	2980
45	996	62	38	3350	3000	2910
46	996	62	38	3340	2980	2910
47	996	62	36	3330	3000	2930
48	996	62	39	3340	2990	2960
49	996	62	38	3330	3000	2890
50	996	62	38	3340	3000	2950
51	1562	174	36	3420	2950	2940
52	1562	160	32	3510	3050	3090
53	1562	128	30	3660	3000	3270
54	1562	96	38	3640	2980	3240
55	1562	88	37	3670	2990	3260
56	1562	82	36	3670	3000	3230
57	1562	82	36	3680	3000	3250
58	1562	82	34	3690	3010	3290
59	1562	82	36	3670	2990	3250
60	1562	82	36	3680	3680	3240
61	498	72	34	4210	4210	3700
62	498	64	32	4560	4560	3940
63	498					
64	498					
65	498					
66	596	48	39	5290	5010	4480
67	596	28	26	5080	5080	4000
68	596	32	36	5340	4980	4640
69	596	48	36	5320	5010	4630
70	596	32	40	5300	5000	4610
71	596	32	40	5310	4980	4590
72	596	28	24	5130	5130	4210
73	596	28	26	5170	4980	4130
74	596	32	40	5280	5010	4400
75	596	32	38	5300	5000	4560

Table A1: Raw Data for Biokinetic Studies (Cont.)

TIME, Days	INFLUENT COD, mg/l	EFFLUENT COD, mg/l	PERMEATE VOLUME, l	TANK MLSS, mg/l		TANK MLVSS, mg/l
				Before Wasting	After Wasting	
77	596	32	40	5290	4950	4630
76	596	32	40	5300	4980	4620
78	996	96	38	5420	5040	4660
79	996	84	34	5520	5000	4870
80	996	64	32	5580	5030	4840
81	996	64	38	5620	5010	4780
82	996	48	37	5650	5000	4860
83	996	48	36	5660	5010	4860
84	996	48	35	5650	4980	4920
85	996	48	36	5640	5000	4860
86	996	48	36	5650	5020	4880
87	996	48	36	5660	4940	4860
88	1486	160	34	5780	5100	4900
89	1486	128	32	5850	4950	4800
90	1486	96	30	5900	5040	4920
91	1486	80	36	6010	5000	5050
92	1486	80	35	5980	4980	5020
93	1486	80	35	5970	5020	5090
94	1486	48	20	5340	5000	4210
95	1486	80	35	5980	4990	5000
96	1486	80	35	5970	5000	5030
97	1992	174	32	6040	5060	5070
98	1992	160	30	6110	4960	5100
99	1992	92	24	5650	5080	4340
100	1992	128	34	6030	4980	4930
101	1992	112	33	6070	4990	5010
102	1992	96	32	6140	5000	5120
103	1992	96	32	6150	5010	5170
104	1992	96	31	6170	4980	5160
105	1992	96	33	6130	5000	5160
106	1992	96	32	6150	4990	5160
107	1992	96	32	6150	5000	5120
108	3014	214	30	6310	4930	5180
109	3014	198	32	6490	4960	5240
110	3014	174	31	6650	4900	5530
111	3014	160	32	6650	4980	5510
112	3014	128	30	6800	5000	5580
113	3014	128	30	6820	4970	5620
114	3014	128	30	6780	5010	5670

Table A1: Raw Data for Biokinetic Studies (Cont.)

TIME, Days	INFLUENT COD, mg/l	EFFLUENT COD, mg/l	PERMEATE VOLUME, l	TANK MLSS, mg/l		TANK MLVSS, mg/l
				Before Wasting	After Wasting	
115	3014	112	26	6110	4960	4810
116	3014	128	31	6670	4990	5460
117	3014	128	30	6800	5000	5610
118	3014	128	30	6810	4960	5620
119	996	116	34	5810	4100	4750
120	996	102	32	4520	3150	3870
121	2246	158	30	3530	3000	3100
122	2246	98	14	3760	3050	3050
123	2246	116	35	3870	3100	3410
124	2246	48	22	3480	3090	2880
125	2246	102	36	3860	2980	3310
126	2246	116	35	3930	3010	3440
127	2246	102	34	3960	3000	3510
128	2246	102	34	3960	2980	3500
129	2246	102	34	3950	3000	3500
130	2246	102	33	3960	3020	3480
131	2994	174	32	4010	3010	3520
132	2994	62	25	3730	2960	3010
133	2994	152	34	4280	3030	3700
134	2994	148	33	4410	2990	3920
135	2994	148	34	4400	3010	3900
136	2994	132	32	4420	2980	3970
137	2994	132	32	4400	3000	3960
138	2994	132	31	4420	3010	4000
139	2994	132	32	4420	4420	3970
140	996	88	30	5010	5010	4340
141	996	74	34	5430	5430	4440
142	1458	162	33	5890	5600	4830
143	1458	156	32	6470	6320	5160
144	1458	156	30	7320	7180	5710
145	1458	68	20	7640	7640	5400
146	1458	96	36	8280	8280	6350
147	1458	96	36	9130	9130	6870
148	1458	96	34	10050	9800	7430
149	996	82	33	10220	9960	7850
150	574	62	23	10150	9950	7040
151	574	54	38	10290	10000	7570
152	574	32	36	10310	9940	7980
153	574	32	36	10280	9980	8030

Table A1: Raw Data for Biokinetic Studies (Cont.)

TIME, Days	INFLUENT COD, mg/l	EFFLUENT COD, mg/l	PERMEATE VOLUME, l	TANK MLSS, mg/l		TANK MLVSS, mg/l
				Before Wasting	After Wasting	
154	574	26	38	10310	10000	8030
155	574	26	38	10320	9990	7950
156	574	26	36	10320	9990	8070
157	574	26	38	10310	9970	8000
158	1168	74	37	10450	10030	7950
159	1168	42	22	10280	9970	7250
160	1168	42	37	10660	10020	7950
161	1168	36	36	10750	10000	8270
162	1168	36	36	10750	9970	8200
163	1168	28	24	10560	10000	7240
164	1168	36	36	10750	10020	8050
165	2086	38	26	10610	9970	7330
166	2086	96	34	10980	10030	8090
167	2086	96	37	11360	9990	8590
168	2086	62	36	11460	10000	8810
169	2086	62	35	11470	10020	8770
170	2086	62	36	11480	9990	8700
171	2086	62	36	11470	9890	8730
172	3112	156	34	11980	10020	9180
173	3112	134	33	12210	9990	9280
174	3112	126	32	12250	9980	9200
175	3112	118	34	12240	10000	9250
176	3112	118	34	12260	9990	9200
177	3112	118	34	12240	10000	9230
178	3112	118	34	12260	10040	9290
179	3984	212	32	12470	9980	9360
180	3984	118	26	11630	9980	7970
181	3984	156	34	12660	10050	9290
182	3984	144	32	12930	9980	9670
183	3984	132	34	12970	9990	9740
184	3984	132	34	12970	10000	9710
185	3984	132	34	12980	10010	9760
186	3984	132	34	12980	10000	9720
187	996	132	34	13020	10020	9600
188	996	26	36	10820	10500	7980
189	996	26	36	11240	10820	8120
190	996	26	36	11400	11000	8540
191	1458	26	34	12110	11750	9080
192	1458	26	34	12760	12250	9250

Table A1: Raw Data for Biokinetic Studies (Cont.)

TIME, Days	INFLUENT COD, mg/l	EFFLUENT COD, mg/l	PERMEATE VOLUME, l	TANK MLSS, mg/l		TANK MLVSS, mg/l
				Before Wasting	After Wasting	
193	1458	26	36	13200	12880	9800
194	1458	26	36	13840	13550	9990
195	1458	26	36	14230	13990	10290
196	1458	18	28	14190	14190	9940
197	1458	26	36	14800	14520	10640
198	1458	26	36	14950	14640	10930
199	1458	26	34	15100	14640	11210
200	926	26	34	15420	14820	11440
201	926	18	36	15640	15040	11550
202	926	18	36	15620	14980	11480
203	926	18	36	15680	15050	11500
204	926	18	36	15650	15010	11560
205	926	18	36	15670	14980	11520
206	926	18	36	15660	15000	11580
207	926	18	36	15660	14990	11540
208	926	18	36	15660	15020	11490
209	1988	18	36	15680	14970	11550
210	1988	80	34	15820	15000	11680
211	1988	72	34	15990	15120	11780
212	1988	72	36	16110	15050	11810
213	1988	48	36	16250	14990	11950
214	1988	48	35	16290	15010	11970
215	1988	48	36	16280	15020	11960
216	1988	48	36	16290	14990	11960
217	1988	48	36	16280	15000	11940
218	1988	48	36	16270	15020	11960
219	1988	48	36	16280	14980	11950
220	2642	48	36	16320	15020	11450
221	2642	182	35	16550	15090	12140
222	2642	146	35	17010	15000	12600
223	2642	98	34	17140	15030	12610
224	2642	92	34	17150	15010	12640
225	2642	86	34	17140	15000	12620
226	2642	86	34	17150	15030	12600
227	2642	86	34	17140	14980	12610
228	2642	86	34	17130	15010	12600
229	2642	86	34	17140	15000	12590
230	3246	86	34	17200	15050	12630
231	3246	224	32	17290	14960	12160

Table A1: Raw Data for Biokinetic Studies (Cont.)

TIME, Days	INFLUENT COD, mg/l	EFFLUENT COD, mg/l	PERMEATE VOLUME, l	TANK MLSS, mg/l		TANK MLVSS, mg/l
				Before Wasting	After Wasting	
232	3246	188	38	17450	14950	12700
233	3246	152	36	17550	15010	12850
234	3246	136	34	17510	14990	12840
235	3246	98	34	17520	15010	12850
236	3246	98	34	17540	14980	12840
237	3246	98	34	17500	15000	12820
238	3246	98	34	17520	15010	12830
239	3246	98	34	17520	14990	12840
240	4468	98	32	17790	15120	12890
241	4468	282	37	18240	15020	13250
242	4468	174	35	18560	15220	13570
243	4468	138	30	18100	15060	12860
244	4468	138	36	18620	15020	13450
245	4468	126	36	18640	15000	13600
246	4468	126	34	18650	15030	13590
247	4468	126	34	18650	15000	13600
248	4468	126	34	18640	14980	13580
249	4468	126	34	18650	15020	13600
250	4468	126	34	18650	15000	13600

APPENDIX B

RAW DATA FOR MICROBIOLOGICAL STUDIES

Table B1: Raw data for Plate Count Results

Tank MLSS Mg/l	Plate Count in the Tank PFU/100 ml	Plate Count in the Permeate PFU/100 ml
3400	1.4E+09	4.6E+06
5000	5.1E+09	4.4E+07
10450	1.1E+10	5.9E+07
10800	3.6E+10	3.6E+07
15000	5.9E+11	5.0E+07

Table B2: Raw Data for Variation of Total Coliform

Tank MLSS Mg/l	Tank Coliform MPN/100 ml	Permeate Coliform MPN/100 ml
3400	1.7E+11	1.7E+07
5000	9.0E+11	5.0E+07
10450	8.0E+11	5.0E+07
10800	7.0E+11	9.0E+07
15000	1.3E+13	2.0E+07

Table B3: Raw Data for Variation of Feacal Coliform

Tank MLSS Mg/l	Tank Feacal Coliform MPN/100 ml	Permeate Coliform MPN/100 ml
3400	7.1E+10	1.4E+06
5000	9.2E+10	1.8E+06
10450	4.0E+11	2.7E+07
10800	4.0E+11	3.4E+07
15000	1.1E+13	2.0E+06

Table B4: Variation of Feacal Streptococcus

Tank MLSS Mg/l	Tank Feacal Streptococcus MPN/100 ml	Permeate Streptococcus MPN/100 ml
3400	1.3E+10	9.0E+05
5000	2.8E+11	9.0E+05
10450	2.0E+11	5.0E+06
10800	4.0E+11	5.0E+06
15000	1.1E+13	2.6E+06

APPENDIX C

RAW DATA FOR SHOCK LOADING AND TOXIC LOADING

Table C1: Raw Data for Shock Loading (High OLR)

Time Days	Influent COD Mg/l	Effluent COD Mg/l	PermeateVolume L	Tank MLSS Mg/l
251	4998	132	38	18460
252	8024	254	36	20040
253	8024	172	34	20850
254	12100	328	38	22340
255	12100	184	36	23180
256	16000	386	38	24940
257	16000	202	36	26190
258	1998	32	38	18450
251	4998	132	38	18460

Table C2: Data for Phenolic Loading Without Acclimatization

Time Days	Influent COD Mg/l	Effluent COD Mg/l	Influent Phenol Mg/l	Effluent Phenol Mg/l
0.0	1000		400	
0.6	1000	910	400	345
0.8	1000	901	400	328
1.0	1000	850	400	305
1.2	1000	821	400	298
1.6	1000	800	400	286
1.9	1000	742	400	275
2.2	1000	710	400	251
2.5	1000	690	400	238
3.0	1000	650	400	225
3.3	1000	592	400	208.3
3.7	1000	574	400	205
4.0	1000	536	400	198
5.0	1000	498	400	186
6.0	1000	459	400	178
7.0	1000	412	400	164
8.0	1000	358	400	160
9.0	1000	312	400	152
10.0	1000	274	400	144
11.0	1000	246	400	136
12.0	1000	228	400	126
13.0	1000	224	400	120
14.0	1000	224	400	120
15.0	1000	224	400	119

Table C3: Raw Data for Phenolic Loading (With Acclimatization)

Time Days	Influent OD (Mg/l)	Effluent COD Mg/l	Influent Phenol Mg/l	Effluent Phenol Mg/l
285	998	126	50	28.4
286	998	68	50	17.9
287	998	57	50	12.6
288	998	48	50	4.1
289	998	42	50	4.1
290	998	42	50	3.8
291	998	38	50	2.4
292	998	34	50	1.2
293	998	34	50	0.8
294	998	34	50	0.5
295	998	34	75	0.4
296	998	148	100	42.1
297	998	136	100	39.5
298	998	124	100	31.4
299	998	108	100	23.8
300	998	92	100	12.5
301	998	78	100	8.7
302	998	56	100	4.6
303	998	48	100	3.9
304	998	42	100	3.2

Table C3: Raw Data for Phenolic Loading (With Acclimatization)(Cont.)

Time Days	Influent OD (Mg/l)	Effluent COD Mg/l	Influent Phenol Mg/l	Effluent Phenol Mg/l
305	998	42	100	2.8
306	998	42	100	2.5
307	998	42	100	2.3
308	998	42	100	2.1
309	998	42	100	2.0
310	998	42	100	2.0
311	998	42	100	2.0
312	998	36	100	1.9
313	998	36	100	1.7
314	998	36	100	1.7
315	998	36	100	1.7
316	998	556	400	282.4
317	998	412	400	218.3
318	998	336	400	156.7
319	998	272	400	112.4
320	998	184	400	98.5
321	998	896	800	598.5
322	998	748	800	524.2
323	998	624	800	487.4
324	998	596	800	401.9
325	998	124		
326	998	32		

Table C4: Raw Data for Chromium Loading

Time Days	Influent COD Mg/l	Effluent COD Mg/l	Influent Chromium Mg/l	Effluent Chromium Mg/l
402	998	486	20	16.5
403	998	132	20	16.4
404	998	664	40	34.6
405	998	318	40	32.6
406	998	738	50	42.3
407	998	84		
408	998	32		

APPENDIX D

RAW DATA FOR MEMBRANE FOULING CONTROL

Table D1: Raw Data for Flux and Turbidity with Clean Membrane

Time	Flux, $L.m^{-2}.hr^{-1}$	Turbidity, NTU
0:00		
0:02	744.3	185
0:05	669.1	126
0:10	601.4	74
0:15	541.3	58
0:20	488.7	42
0:30	383.4	24
0:40	318.3	15
0:50	263.1	8.6
1:00	233.1	4.4
1:08	188.0	2.8
1:17	238.1	5.2
1:20	219.3	3.4
1:30	190.5	2.9
1:45	142.8	2.2
2:00	97.7	1.2
2:15	66.4	0.8
2:23	52.6	0.5
2:32	81.4	1.4
2:35	70.2	0.82
2:40	56.4	0.54
2:55	47.6	0.36
3:10	42.6	0.22
3:25	40.1	0.15
3:38	37.6	0.15
3:47	60.1	0.72
3:50	50.1	0.48
3:55	42.6	0.2
4:05	38.8	0.18
4:20	35.1	0.18
4:40	31.3	0.17
4:53	28.8	0.15
5:02	56.4	0.47
5:05	52.6	0.35
5:15	45.1	0.2
5:30	38.8	0.18
5:45	35.1	0.15
6:00	30.1	0.12
6:08	27.6	0.1
6:17	55.1	

Table D1: Raw Data for Flux and Turbidity with Clean Membrane (Cont.)

Time	Flux, $L.m^{-2}.\ hr^{-1}$
6:20	52.6
6:30	45.1
6:45	37.6
7:00	35.1
7:15	30.1
7:23	27.6
7:32	55.1
7:35	51.4
7:45	43.9
8:00	36.3
8:15	33.8
8:30	30.1
8:38	27.6
8:47	55.1
8:50	52.6
9:00	47.6
9:15	40.1
9:30	32.6
9:45	30.1
9:53	27.6
10:02	53.9
10:05	51.4
10:15	43.9
10:30	37.6

Table D1: Raw Data for Flux and Turbidity with Clean Membrane (Cont.)

Time	Flux, $L.m^{-2}.\ hr^{-1}$
10:45	32.6
11:00	30.1
11:08	27.6
11:17	50.1
11:20	43.9
11:30	36.3
11:45	31.3
12:00	28.8
12:15	27.6
12:23	25.1
12:32	169.2
12:35	135.3
12:45	102.7
13:00	72.7
13:15	61.4
13:30	47.6
13:38	40.1
13:47	68.9
14:00	52.6
14:15	40.1
14:30	35.1
14:45	32.6
14:53	30.1
15:02	62.7

Table D1: Raw Data for Flux and Turbidity with Clean Membrane (Cont.)

Time	Flux, $L.m^{-2}.\ hr^{-1}$
15:15	52.6
15:30	45.1
15:45	40.1
16:00	35.1
16:08	28.8
16:17	57.6
16:30	47.6
16:45	42.6
17:00	37.6
17:15	35.1
17:23	28.8
17:32	52.6
17:45	47.6
18:00	42.6
18:15	37.6
18:30	32.6
18:38	27.6
18:47	52.6
19:00	46.4
19:15	42.6
19:30	37.6
19:45	32.6
19:53	27.6
20:02	52.6

Table D1: Raw Data for Flux and Turbidity with Clean Membrane (Cont.)

Time	Flux, $L.m^{-2}.\ hr^{-1}$
20:15	45.1
20:30	40.1
20:45	35.1
21:00	31.3
21:08	27.6
21:17	50.1
21:30	41.3
21:45	35.1
22:00	31.3
22:15	28.8
22:23	27.6
22:32	50.1
22:45	41.3
23:00	36.3
23:15	30.1
23:30	27.6
23:38	25.1

Table D2: Raw Data for Flux and Turbidity during a typical day

Time	Flux, $L.m^{-2}.\ hr^{-1}$	Turbidity, NTU
0:00		
0:02	172.9	2.36
0:05	135.3	0.96
0:10	102.7	0.78
0:15	87.7	0.67
0:20	77.7	0.58
0:30	68.9	0.49
0:40	60.1	0.45
0:50	52.6	0.38
1:00	43.9	0.36
1:08	40.1	0.32
1:17	68.9	1.2
1:20	62.7	0.88
1:30	57.6	0.72
1:45	47.6	0.38
2:00	40.1	0.24
2:15	32.6	0.12
2:23	30.1	0.1
2:32	62.7	0.82

Table D2: Raw Data for Flux and Turbidity during a typical day (Cont.)

Time	Flux, $L.m^{-2}.\ hr^{-1}$	Turbidity, NTU
2:35	56.4	0.74
2:40	52.6	0.48
2:55	48.9	0.4
3:10	40.1	0.24
3:25	32.6	0.16
3:38	30.1	0.12
3:47	60.1	0.75
3:50	50.1	0.48
3:55	42.6	0.39
4:05	38.8	0.24
4:20	35.1	0.15
4:40	31.3	0.11
4:53	28.8	0
5:02	56.4	0.62
5:05	52.6	0.47
5:15	45.1	0.35
5:30	38.8	0.22
5:45	35.1	0.1
6:00	30.1	0
6:08	27.6	0

Table D2: Raw Data for Flux and Turbidity during a typical day (Cont.)

Time	Flux, $L.m^{-2}.\ hr^{-1}$	Turbidity, NTU
6:17	55.1	0.58
6:20	52.6	0.44
6:30	45.1	0.29
6:45	37.6	0.18
7:00	35.1	0
7:15	30.1	0
7:23	27.6	0
7:32	55.1	0.58
7:35	51.4	0.46
7:45	43.9	0.31
8:00	36.3	0.1
8:15	33.8	0
8:30	30.1	0
8:38	27.6	0
8:47	55.1	0.58
8:50	52.6	0.5
9:00	47.6	0.29
9:15	40.1	0.24
9:30	32.6	0.1
9:45	30.1	0

Table D2: Raw Data for Flux and Turbidity during a typical day (Cont.)

Time	Flux, $L.m^{-2}.\ hr^{-1}$	Turbidity, NTU
9:53	27.6	0
10:02	53.9	0.56
10:05	51.4	0.48
10:15	43.9	0.25
10:30	37.6	0.13
10:45	32.6	0
11:00	30.1	0
11:08	27.6	0
11:17	50.1	0.51
11:20	43.9	0.35
11:30	36.3	0.11
11:45	31.3	0
12:00	28.8	0
12:15	27.6	0
12:23	25.1	0
12:32	169.2	2.34
12:35	135.3	1.12
12:45	102.7	0.75
13:00	72.7	0.66
13:15	61.4	0.48

Table D2: Raw Data for Flux and Turbidity during a typical day (Cont.)

Time	Flux, $L.m^{-2}.\ hr^{-1}$	Turbidity, NTU
13:30	47.6	0.42
13:38	40.1	0.38
13:47	68.9	1.18
14:00	52.6	0.62
14:15	40.1	0.36
14:30	35.1	0.25
14:45	32.6	0.13
14:53	30.1	0.1
15:02	62.7	0.58
15:15	52.6	0.44
15:30	45.1	0.36
15:45	40.1	0.24
16:00	35.1	0.12
16:08	28.8	0
16:17	57.6	0.52
16:30	47.6	0.47
16:45	42.6	0.29
17:00	37.6	0.16
17:15	35.1	0.1
17:23	28.8	0

Table D2: Raw Data for Flux and Turbidity during a typical day (Cont.)

Time	Flux, $L.m^{-2}.\ hr^{-1}$	Turbidity, NTU
17:32	52.6	0.44
17:45	47.6	0.32
18:00	42.6	0.26
18:15	37.6	0.15
18:30	32.6	0
18:38	27.6	0
18:47	52.6	0.45
19:00	46.4	0.32
19:15	42.6	0.28
19:30	37.6	0.16
19:45	32.6	0.1
19:53	27.6	0
20:02	52.6	0.44
20:15	45.1	0.28
20:30	40.1	0.16
20:45	35.1	0.11
21:00	31.3	0
21:08	27.6	0
21:17	50.1	0.44
21:30	41.3	0.28

Table D2: Raw Data for Flux and Turbidity during a typical day (Cont.)

Time	Flux, $L.m^{-2}.\ hr^{-1}$	Turbidity, NTU
21:45	35.1	0.19
22:00	31.3	0.1
22:15	28.8	0
22:23	27.6	0
22:32	50.1	0.41
22:45	41.3	0.29
23:00	36.3	0.16
23:15	30.1	0
23:30	27.6	0
23:38	25.1	0

Table D3: Raw Data for Flux and Turbidity after Back Washing

Time Hours	After Back Washing With Air Alone		After Back Washing With Air and Water		After Back Washing With Water Alone	
	Flux, $L.m^{-2}.\ hr^{-1}$	Turbidity, NTU	Flux, $L.m^{-2}.\ hr^{-1}$	Turbidity, NTU	Flux, $L.m^{-2}.\ hr^{-1}$	Turbidity, NTU
0:02	55.1	0.58	55.1	0.58	55.1	0.58
0:05	52.6	0.5	52.6	0.5	52.6	0.5
0:15	47.6	0.29	47.6	0.29	47.6	0.29
0:30	40.1	0.24	40.1	0.24	40.1	0.24
0:45	32.6	0.1	32.6	0.1	32.6	0.1
1:00	30.1	0	30.1	0	30.1	0
1:08	27.6	0	27.6	0	27.6	0
1:17	53.9	0.56	53.9	0.56	53.9	0.56
1:20	51.4	0.48	51.4	0.48	51.4	0.48
1:30	43.9	0.25	43.9	0.25	43.9	0.25
1:45	37.6	0.13	37.6	0.13	37.6	0.13
2:00	32.6	0	32.6	0	32.6	0
2:15	30.1	0	30.1	0	30.1	0
2:23	27.6	0	27.6	0	27.6	0
2:32	50.1	0.51	50.1	0.51	50.1	0.51
2:35	43.9	0.35	43.9	0.35	43.9	0.35
2:45	36.3	0.11	36.3	0.11	36.3	0.11

Table D3: Raw Data for Flux and Turbidity after Back Washing (Cont.)

Time Hours	After Back Washing With Air Alone	After Back Washing With Air and Water	After Back Washing With Water Alone	Time Hours	After Back Washing With Air Alone	After Back Washing With Air and Water
3:15	28.8	0	28.8	0	28.8	0
3:30	27.6	0	27.6	0	27.6	0
3:38	25.1	0	25.1	0	25.1	0
3:47	169.2	2.34	180.4	3.32	150.4	2.22
3:50	135.3	1.12	165.4	1.84	127.8	1.08
4:00	102.7	0.75	137.8	1.22	97.7	0.67
4:15	72.7	0.66	111.5	1.02	68.9	0.59
4:30	61.4	0.48	75.2	0.88	58.9	0.44
4:45	47.6	0.42	60.1	0.53	42.6	0.4
4:53	40.1	0.38	43.9	0.46	40.1	0.38
5:02	68.9	1.18	75.2	1.42	66.4	1.02
5:15	52.6	0.62	60.1	0.74	52.6	0.58
5:30	40.1	0.36	42.6	0.48	40.1	0.34
5:45	35.1	0.25	35.1	0.22	35.1	0.22
6:00	32.6	0.13	32.6	0.12	32.6	0.1
6:08	30.1	0.1	30.1	0	30.1	0
6:17	62.7	0.58	65.2	0.6	62.7	0.44

Table D3: Raw Data for Flux and Turbidity after Back Washing (Cont.)

Time Hours	After Back Washing With Air Alone	After Back Washing With Air and Water	After Back Washing With Water Alone	Time Hours	After Back Washing With Air Alone	After Back Washing With Air and Water
6:30	52.6	0.44	53.9	0.48	52.6	0.36
6:45	45.1	0.36	45.1	0.36	45.1	0.24
7:00	40.1	0.24	40.1	0.25	40.1	0.12
7:15	35.1	0.12	35.1	0.14	35.1	0
7:23	28.8	0	28.8	0	28.8	0
7:32	57.6	0.52	57.6	0.52	57.6	0.44
7:45	47.6	0.47	47.6	0.47	47.6	0.37
8:00	42.6	0.29	42.6	0.29	42.6	0.29
8:15	37.6	0.16	37.6	0.16	37.6	0.16
8:30	35.1	0.1	35.1	0.1	35.1	0
8:38	28.8	0	28.8	0	28.8	0
8:47	52.6	0.44	52.6	0.44	52.6	0.44
9:00	47.6	0.32	47.6	0.32	47.6	0.32
9:15	42.6	0.26	42.6	0.26	42.6	0.26

Table D4: Raw Data for Flux and Turbidity after Cleaning Membrane with Brush

Time	Flux, $L.m^{-2}.\ hr^{-1}$	Turbidity, NTU
0:02	55.1	0.56
0:05	52.6	0.48
0:15	47.6	0.24
0:30	40.1	0.13
0:45	32.6	0
1:00	30.1	0
1:08	27.6	0
1:17	53.9	0.48
1:20	51.4	0.34
1:30	43.9	0.15
1:45	37.6	0.1
2:00	30.1	0
2:15	27.6	0
2:23	25.1	0
2:32	43.9	0.48
2:35	35.1	0.23
2:45	27.6	0.12
3:00	25.1	0
3:15	25.1	0

Table D4: Raw Data for Flux and Turbidity after Cleaning Membrane with Brush (Cont.)

Time	Flux, $L.m^{-2}.\ hr^{-1}$	Turbidity, NTU
3:30	22.6	0
3:38	22.6	0
3:47	169.2	185
3:50	135.3	126
4:00	102.7	74
4:15	72.7	30
4:30	61.4	42
4:45	47.6	0.82
4:53	40.1	0.48
5:02	68.9	1.18
5:15	52.6	0.46
5:30	40.1	0.32
5:45	35.1	0.19
6:00	32.6	0.1
6:08	30.1	0
6:17	62.7	0.46
6:30	52.6	0.29

Table D4: Raw Data for Flux and Turbidity after Cleaning Membrane with Brush (Cont.)

Time	Flux, $L.m^{-2}.\ hr^{-1}$	Turbidity, NTU
6:45	45.1	0.12
7:00	40.1	0
7:15	35.1	0
7:23	28.8	0
7:32	57.6	0.44
7:45	47.6	0.3
8:00	42.6	0.11
8:15	37.6	0
8:30	35.1	0
8:38	28.8	0

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